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# RISK ASSESSMENT OF STORAGE AND TRANSPORT OF LIQUEFIED NATURAL GAS AND LP-GAS

SCIENCE APPLICATIONS, INCORPORATED

PREPARED FOR
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16. Abstracts: A method for assessing the societal risk of transporting liquefied petroleum gas (LPG) and liquefied natural gas (LNG) is described, and is illustrated by application to the transport of LPG by tank truck and LNG by tanker ship in the U.S. Data on past experience and projected future handling of these liquefied gases are used with analysis of flammable plume formation and ignition, and population distributions, to estimate the risks of fatalities from tank truck and tanker ship accidents. From an estimated 52 significant accidents per year with LPG tank trucks at the present truck-associated transportation rate of 20 billion gallons of LPG per year, a fatality rate of 1.2 per year is calculated. For the projected 1980 importation of 33 billion gallons by tanker ship, a fatality rate of 0.4 per year is calculated, using a conservatively high one chance in 20,000 of a significant accident per trip. Comparison with fires and explosions from all causes in the U.S. and Canada leading to 10 or more fatalities shows that these are 100 times more frequent than the predicted frequency of comparable LPG and LNG accidents. Tabulations of experience with spills of flammable volatile liquids are included.

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# RISK ASSESSMENT OF STORAGE AND TRANSPORT OF LIQUEFIED NATURAL GAS AND LP-GAS

John A. Simmons

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Final Report for Contract 68.01.2695
Sponsored by the Environmental Protection Agency

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## 1.0 Summary

This report presents the results of a study to develop a method for assessing the societal risk of transporting LP-Gas and LNG (liquefied natural gas). The method is illustrated through its application to the transport of LP-Gas by tank truck and LNG by tankship in the U.S. The measure of risk used was the probability of fatalities resulting from accidental leaks and tank ruptures. Fatalities are a direct, unambiguous measure that facilitates comparison with the risk of other activities via available, historical accident data (e.g. fires, explosions, etc.).

The reason for considering both LNG and LP-Gas in this study is the similarity of their hazards and the large volume transported (or planned to be transported). Both are highly volatile liquids such that spills may create extensive flammable plumes. The plumes are negatively bouyant and tend to lie flat on the ground or water. On the other hand, there are important differences. LNG is a cryogenic liquid (B.P. 113°K), consisting primarily of methane, and is stored at atmospheric pressure. LPG consists primarily of liquid propane and is stored at ambient temperatures at a pressure of about 100 psig. Because of this, a spill results in the immediate flashing of 30 to 35 percent into vapor.

In 1972 $20 \times 10^9$ gal of LP-Gas was transported by truck or a combination of pipeline and truck. The average truckload is 4,370 gal. By 1980 the amount is expected to rise to $25$ to $29 \times 10^9$ gal. It is projected that in 1980 the U.S. will be importing $33 \times 10^9$ gal of LNG ($2.06 \times 10^{12}$ ft$^3$ of natural gas) in tankships carrying approximately $32.5 \times 10^6$ gal each.

The only known major accident involving LNG was the collapse of a $1.6 \times 10^6$ gal "peakshaving" storage tank in Cleveland, Ohio, in 1944.[18] The resulting fires and explosions killed 135 people and did $10,000,000 property damage. The tank collapse was caused by the low temperature brittle fracture of the tank wall. A major contributing factor was that the tank was not diked to prevent the spread of the burning liquid. This type of accident is now unlikely, since storage tanks are diked and proven materials of construction are available for cryogenic service.

Examination of accounts of accidents involving spills of LP-Gas and other volatile fuels, indicated that the major fraction of deaths, injiries and property damage is caused by a flash fire in a large flammable vapor plume formed prior to ignition. In some cases explosions and even detonations may result when a portion of the plume has infiltrated a building or other confined region. Other types of fires, such as liquid pool fires, often cause extensive property damage but only infrequently cause fatalities and injuries. The explosive rupture of LP-Gas tanks, because of overheating in a fire, is another mechanism which causes many fatalities. However, this type of accident is unlikely for LNG, because its storage vessels are not of a type permitting the buildup of high pressure in a fire. For these reasons, the risk assessment methodology was developed primarily for accidents which lead to the formation of a flammable plume.

The risk assessment method is based on an event tree model of the sequence of occurrences following an accident. There are four events: (1) the spill and generation of vapor at a given rate; (2) dispersal and mixing of the vapor with air; (3) ignition and burning of the flammable mixture;

(4) death of all persons within the flammable plume. Each of the events is characterized quantitatively by discretized variables that have one or more values. Specification of a value for each variable specifies a single accident event chain with a definite number of fatalities, ρA (population density x ground area covered by the flammable plume at the time of ignition). The frequency of occurrence of that consequence is given by the product of the probabilities of the values for each variable.

For LP-Gas tank trucks spill size and frequency was based on reviews of recent accident experience, which indicated an accident and tank rupture frequency of 14 per year. There is no direct accident experience for LNG tankships. Their design is new and is believed to be very crash resistant. The cargo tanks ($\sim$ 25,000 $m^3$) are mounted inside a double hull, which are separated by several feet and are tied together by web frames, stringers, etc. Because of the lack of direct data it was decided to calculate an upper limit for the risk of LNG tanker spills. Consequently the spill frequency was assumed to be the same as was found for conventional oil tankers, by which the rupture of a 25,000 $m^3$ LNG cargo tank is expected $2 \times 10^{-5}$ times per trip.

A Gaussian plume model was used to describe the mixing and dispersion of the vapor. Pasquill atmospheric stability parameters, modified to account for the negative buoyancy of the plume, were used. The joint frequency of occurrence of wind speed and a Pasquill stability category were derived from U.S. Weather Bureau data.

The plume model together with the lower flammability limit permits estimation of the maximum ground area that the flammable plume may cover. That the plume will ever achieve such a size is unlikely because of the many ignition sources in populated areas. For a region comprising a given population density and type of activity (industrial, commercial, etc.), it is thought reasonable to suppose that the probability of plume ignition should depend only on the area of the flammable portion. Data on flash fires involving vapor plumes of volatile fuels for the years 1930 to 1974 were reviewed and of these data the plume area at the time of ignition could be estimated for 57 incidents. The result was that the probability of ignition is distributed normally with the logarithm of the flammable area. Fifty percent of the plumes are ignited almost immediately, area < 30 $m^2$. The derived relationship predicts a probability of 0.004 that the maximum area plume, ∿ 0.6 sq. mi., will be attained following a spill of 25,000 $m^3$ of LNG on water.

Population density frequency distribution appropriate to LP-Gas tank truck spills was the same as used in the similar risk assessment for chlorine rail tank car spills.[1] The density frequency distribution used for the LNG spills was that for the land areas adjacent to Raritan Bay and the Arthur Kill in New York Harbor.

The statistical number of fatalities and frequency of occurrence were computed for each combination of values for the discretized variables, and the frequency of fatalities within evenly divided ranges were summed. Summing the products of frequency and statistical fatalities for all ranges yields the expected annual average fatalities per year. For LNG tankship accidents assuming 1,000 annual trips, this is 0.4

fatalities/year; for the LP-Gas tank truck accidents it is 1.2 fatalities per year. The latter number agrees surprisingly well with the value of 1 to 2 fatalities per year derived from accident data for the years 1931 to 1974.

Comparison with large fatality fires and explosions (fatalities > 10) from all causes in the U.S. and Canada shows that these are 100 times more frequent than the predicted frequency of comparable LNG and LP-Gas accidents.

There are too few historical data to verify the predictions of frequency of accidents with a given number of fatalities. Nevertheless, based on the quality of the data used to develop the frequency values for the four variables, the overall uncertainty for the LP-Gas spills was estimated to be a factor of 10 to 1/10. For LNG tanker spills an uncertainty of a factor of 2 to 1/100 was esimated. The latter factor primarily reflects the disregard of the protection afforded by the double hull design.

## 2.0 Introduction

The objective of this study was to demonstrate a methodology for assessing the societal risk of transporting LP-Gas and LNG (liquefied natural gas). Here societal risk means the possibility of undesirable consequences to the general population caused by leaks or spills of these materials. The entire spectrum of consequences, ranging from the smallest to the largest together with the likelihood of occurrence of each level of consequence, make up the valuation of risk. Presentation may be either a graph or histogram of consequence versus frequency or the sum of the products of frequency and consequence, which is the annual average level of consequence.

In principle, consequence might refer to one or more environmental or public safety factors. For transport of LP-Gas and LNG the primary risk is to public safety because of the severe fire and explosion hazards, and for this a direct measure of consequence is the number of fatalities resulting from an accident. This measure also is convenient because it facilitates comparison with societal risks of other activities through the large body of historical accident mortality data.

The methodology described here is similar to that used to assess the risk of the transport of liquid chlorine.[1] To illustrate the methodology in that study the expected number of fatalities from chlorine rail tank car accidents were estimated. Here we estimated the fatalities from LP-Gas tank truck accidents and LNG tankship collisions and groundings in U.S. harbors near planned (by 1980) and existing LNG importation terminals.

The reason for considering both LNG and LP-Gas in this study is that their hazard characteristics are similar. Both liquids when spilled form a negatively buoyant fuel-air cloud which may be flammable and in some cases detonable.[2] (The detonation of LP-Gas vapor-air clouds has occurred in several accidents; detonation of LNG vapor-air clouds under severe accident conditions is conceivable but has not been reported.) According to the Committee on Hazardous Materials, methane, the major constituent of LNG, and propane, the major constituent of LP-Gas, have been given a maximum fire hazard rating of 4 and minimum rating of zero as poisons or pollutants.[3]

However, there are differences, especially in the manner the liquids are transported and stored. LNG is a cryogenic fluid boiling at approximately -260°F, depending on its composition. LP-Gas is a liquid stored at ambient temperatures at which its vapor pressure is approximately 100 psia. Vessels containing LNG are well insulated and are vented in a manner to prevent overpressures of more than a few psig. LP-Gas vessels generally are uninsulated pressure vessels with pressure relief valves set to open at 200 to 300 psig.

According to a recent description,[4] DISTRIGAS Staten Island LNG terminal by July 1974 will have the capability of storing $1.8 \times 10^6$ barrels (286,000 $m^3$) of LNG ($6.3 \times 10^9$ $ft^3$ of natural gas) in two concrete tanks. By 1976 they expect to be supplying $750 \times 10^6$ $ft^3$ of NG per day into the gas distribution system. The tanks will be supplied every two to three days by a LNG tanker carrying 775,000 barrels (125,000 $m^3$) of LNG ($2.6 \times 10^9$ $ft^3$ NG).

The DISTRIGAS terminal is one of the smaller planned. Table I lists the LNG importation terminals for which approval by the Federal Power Commission has been granted or applied for and which are to be in operation by 1980.[5] With all of these terminals in operation the number of tankship trips (125,000 $m^3$ of LNG per trip), needed to make the deliveries indicated, total 999. To date only a few hundred thousand cubic meters of LNG have been actually imported into the U.S. (which was at the Everett, Massachusetts, terminal).

Besides these importation terminals there exist (December, 1972) 70 liquefaction plants and/or storage facilities, used primarily for "peak-shaving."[6] The capacity of the storage tanks at these facilities range from 250 to $5.75 \times 10^5$ bbls. Currently, the storage facilities without a liquefaction plant (termed a "LNG Satellite Installation") are supplied by tank trucks carrying 10,000 to 11,000 gal (238 to 262 bbls). Additionally, transport by barge and rail are being considered. In particular, DISTRIGAS plans to transport LNG via 30,000-barrel barges from its Staten Island terminal to terminals in Brooklyn and Queens, New York.[4]

LP-Gas generally is stored in bulk in smaller tanks than LNG; these range in capacity from 2,000 to over 30,000 gal (48 to over 720 bbls). The exact number of such tanks within the U.S. has not been determined. Transportation is via tank truck (up to 12,000 gal capacity), railroad tank car (10,000 to over 32,000 gal capacity), barge and pipeline. Table II gives the amount of LP-Gas shipped by the various modes for the years 1969 through 1972.[7] From this table it is evident that the total amount of LP-Gas transported has been increasing at an average rate of approximately 5 percent per

TABLE I: LNG IMPORT TERMINAL PROJECTS[a]

| Terminal | Importer | Date Planned to Begin Operation | Delivery ($10^9$ ft$^3$/year) | Annual Number of Tankship Trips[c] |
|---|---|---|---|---|
| Everett, Mass. & Staten Island, N.Y. | Distrigas Corp. | 1971, 1975 | 60[b] | 22 |
| Cove Point, Md. | Columbia LNG Corp. | 1976-1977 | 116 | 43 |
| | Consolidated System LNG Co. | 1976-1977 | 198[b] | 73 |
| Staten Island, N.Y. and Providence, R.I. | Eascogas LNG, Inc. | 1975-1976 | 238[b] | 87 |
| Gloucester Co., N.J. | El Paso Eastern Co. | 1978 | 154 | 57 |
| | Transco Energy Co. | 1978 | 154 | 57 |
| Savannah, Ga. | Southern Energy Co. | 1976-1978 | 177[b] | 65 |
| Lake Charles, La. | Trunkline LNG Co. | 1979 | 179 | 66 |
| Port Hueneme, Ca. | Pacific Indonesia LNG Co. | 1979 | 226 | 83 |
| Port Concepcion, Ca. | El Paso Alaska | 1980 | 1,220 | 446 |

[a] Imports applied for to the Federal Power Commission to begin by 1980.

[b] Maximum amount applied for; the amounts vary over the years.

[c] Calculated assuming tankships of 125,000 m$^3$ capacity.

## TABLE II - LP-GAS TRANSPORTATION

### ANNUAL SHIPMENTS, MILLIONS OF GALLONS

| Mode | 1969 | | 1970 | | 1971 | | 1972 | |
|---|---|---|---|---|---|---|---|---|
| | $10^6$gal. | Per Cent | $10^6$gal. | Per Cent | $10^6$gal. | Per Cent | $10^6$gal. | Per Cent |
| Truck | 4,546 | 24 | 3,423 | 18 | 3,713 | 19 | 3,965 | 18 |
| Pipeline-Truck | 11,411 | 61 | 13,049 | 69 | 13,778 | 72 | 16,163 | 74 |
| Rail | 1,857 | 10 | 1,611 | 9 | 1,231 | 6 | 1,171 | 5 |
| Pipeline-Rail | 728 | 4 | 544 | 3 | 368 | 2 | 431 | 2 |
| Tanker or Barge | 174 | 1 | 163 | 1 | 93 | 1 | 103 | 1 |
| TOTAL | 18,716 | 100 | 18,789 | 100 | 19,184 | 100 | 21,834 | 100 |

year. The share transported by truck and the pipeline-truck* combination also has been increasing and now these modes account for more than 90 percent of LP-Gas shipments. The share transported by rail tank car is small by comparison and has been decreasing over the years.

In 1972, approximately 19 x $10^9$ gal of LP-Gas were transported by truck or the pipeline truck combination. Based on projections of U.S. demand of 30 to 35 x $10^9$ gal, it is estimated that 25 to 29 x $10^9$ gal will be transported by these modes in 1980.[30] This amount is comparable to the projected importation of LNG, 5.59 x $10^7$ $m^3$ or 14.6 x $10^9$ gal in 1980.

For purposes of illustrating the risk assessment methodology, the study focused on LNG tankship accidents and LP-Gas tank truck accidents. Time did not permit assessment of the risk of all transport activities and associated accidents. However there are reasons for believing that these accidents are at least major contributors to the risk of transporting these fuels. Without any detailed analysis it was felt that major LNG spills most likely will result from collisions and groundings of these vessels. Typical LNG terminal storage tanks are substantial structures and are not exposed to so many moving, heavy objects as are tankships. Unloading operations usually are spill-prone but the quantities spilled generally are less than those from tank rupture.

Similarly for LP-Gas, the tank truck is vulnerable to puncture via the many forms of vehicular accidents on highways. Spills resulting from accidents during transfer of LP-Gas

*LPG is shipped by pipeline to central distribution points from whence it is distributed further by tank trucks.

to-and-from the tank truck also are prevalent and are considered in the study. Pipeline accidents (e.g. rupture) have occurred but only a few have caused casualties. They were not considered. Rail tank car accidents also were not considered. There have been many spectacular accidents resulting from the puncture of rail tank cars. However, the above data shows that this mode of LP-Gas transport is declining and is being replaced by pipeline and truck transport.

## 3.0 Accidents and Fatality Mechanisms

In this section, we discuss the sequence of events and the physical processes that may occur following a spill of LNG or LP-Gas and that produce consequences such as death, injury and property damage. The discussion indicates that the major fatality-producing mechanism for the general population is the flash fire in vapor-air plumes formed by the spilled liquids. This justifies the emphasis of this analysis on such fires.

### 3.1 Event Sequence Following Spills and Leaks

The possible chains of events and consequences following a spill of LNG or LP-Gas are diagramed in Figure 1. Spilled liquid LNG is cold, and contact with the surrounding ground or water causes the liquid to boil, the rate of vaporization being controlled primarily by the rate of heat transfer from from the water or ground surface. The water and ground are cooled rapidly underneath the LNG and in the case of a water spill, ice may form. However, as the LNG spreads out it continually contacts more warm water or ground. Experiments with up to 2700-gallon spills of LNG onto water indicate that the pool of spilled LNG grows linearly with time, eventually reaching a maximum diameter which varies with the 1/3 power of the amount spilled.[8][9][10] The vaporization rate increases as the pool spreads and the total time required for vaporization also varies with the 1/ 3 power of the amount spilled. Consider the "instantaneous" rupture of a LNG tank of a tankship, 25,000 $m^3$ or $0.157 \times 10^6$ bbls. Predictions based on these data are that the vaporization time will be 710 sec. and the maximum LNG pool diameter will be 1570 ft.[9]

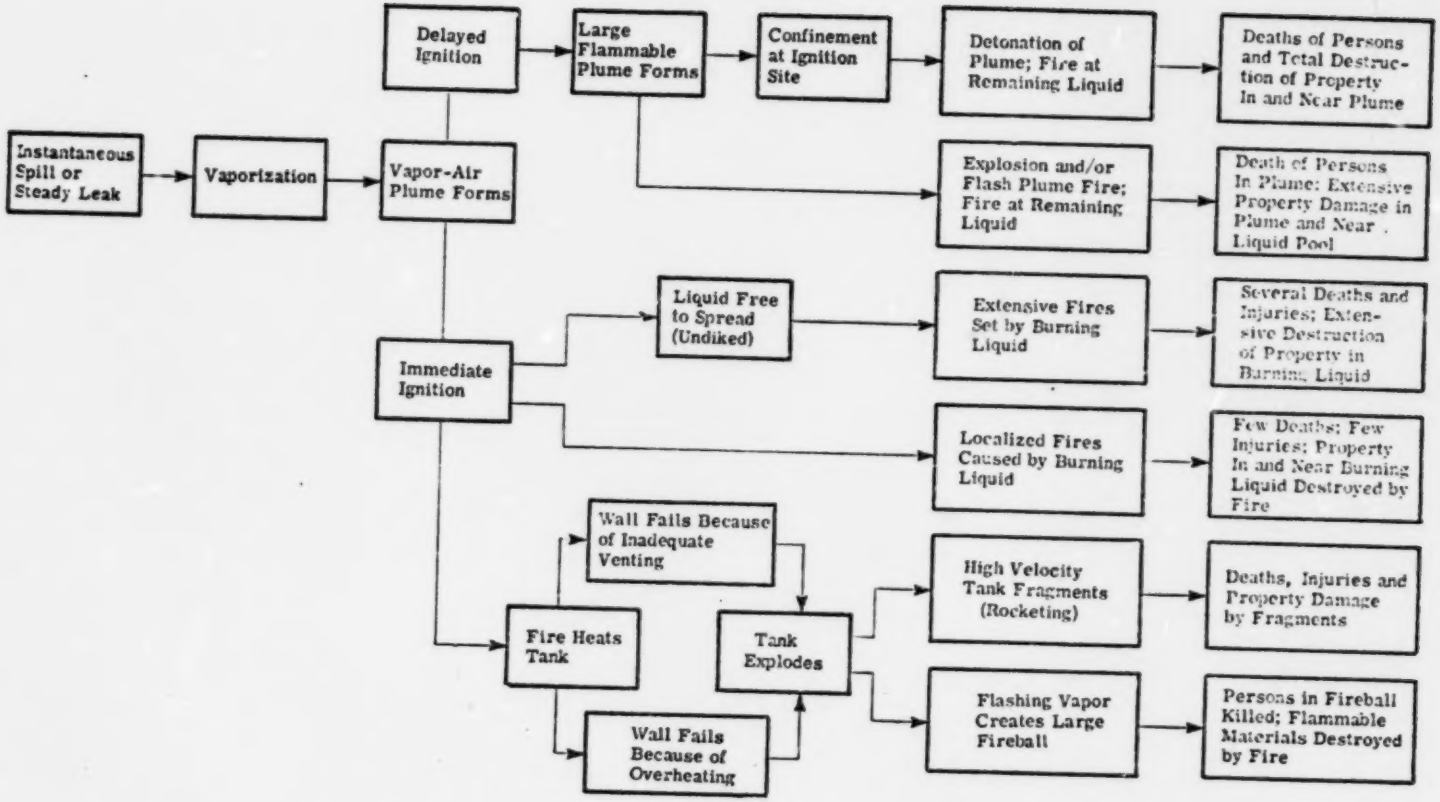
Instantaneous Spill or Steady Leak
Vaporization
Vapor-Air Plume Forms
Delayed Ignition
Large Flammable Plume Forms
Confinement at Ignition Site
Detonation of Plume; Fire at Remaining Liquid
Deaths of Persons and Total Destruction of Property In and Near Plume
Explosion and/or Flash Plume Fire; Fire at Remaining Liquid
Death of Persons In Plume; Extensive Property Damage in Plume and Near Liquid Pool
Immediate Ignition
Liquid Free to Spread (Undiked)
Extensive Fires Set by Burning Liquid
Several Deaths and Injuries; Extensive Destruction of Property in Burning Liquid
Localized Fires Caused by Burning Liquid
Few Deaths; Few Injuries; Property In and Near Burning Liquid Destroyed by Fire
Wall Fails Because of Inadequate Venting
Fire Heats Tank
Tank Explodes
High Velocity Tank Fragments (Rocketing)
Deaths, Injuries and Property Damage by Fragments
Wall Fails Because of Overheating
Flashing Vapor Creates Large Fireball
Persons in Fireball Killed; Flammable Materials Destroyed by Fire

FIGURE 1 - EVENT CHAINS FOR LNG AND LP-GAS LEAKS AND SPILLS

The vapor (primarily methane) from the spill initially is at the temperature of the LNG itself and at this temperature it is 25 to 35 percent more dense than the ambient air (depending on temperature and relative humidity). The vapor mixes rapidly with air (mostly by turbulent diffusion), and because the mixing is nearly adiabatic, the mixtures also are colder than the ambient air.

Calculations show that an adiabaticly mixed LNG vapor-air plume is always denser than ambient air except at high relative humidities (>60%) and when the methane concentration is in the range of 5 to 25 percent.[9] In this range, the plume is only very slightly positively buoyant because of the heat liberated by the condensation of water. At higher dilutions (to 5% methane and less), the plume is warmer, the water droplets evaporate, and the plume returns to its negatively buoyant condition. Calculations in Reference 9 indicated that the mixing is not quite adiabatic because of convective heating from the water surface. However this effect is small for the size of spills they considered.

Experiments show that the plume always lies low on the water surface, implying negative buoyancy at all vapor dilutions.[9][10] The reason for this is suggested by some laboratory experiments in which it was shown that the film boiling of LNG on water is sufficiently violent to eject cold water droplets into the plume (and by implication, droplets of LNG as well).[11] Both of these effects would cause increased cooling and greater negative buoyancy of the plume.

As the vapor and air mix three regions are formed, being classified by the upper and lower flammability limits in air, 15 and 5 volume percent methane, respectively. Nearest the spill site a fuel rich mixture exists. At the periphery of the plume the mixture is too lean and is non-flammable. In between these is a flammable region. The typical shapes of these regions are illustrated in Figure 2. Both their shapes and size depend on the vaporization rate, meteorological conditions and topography. For the LNG tanker spill example, the flammable portion of the plume was predicted to extend approximately 6 miles downwind under neutral atmospheric conditions and a 24 mph wind speed.[9]

However, implicit in this prediction is that the plume remains unignited. When ignited, the flammable portion burns immediately. The fuel rich portion burns somewhat more slowly as a very turbulent diffusion flame. After the plume is consumed, the fire seats on the vaporizing pool of remaining liquid. If the plume is not ignited, mixing continues and the vapor eventually becomes diluted to a concentration below the lower flammable limit.

Spills and leaks of LP-Gas behave somewhat differently since the liquid is stored under pressure. When released, it first flashes adiabaticly, in which the amount of vapor evolved is sufficient to cool the remaining liquid to -43.7°F, at which temperature the vapor pressure is 1 atm. In this way a large vapor cloud forms immediately; for example, if the initial temperature is 68°F, 31 percent of the liquid flashes to vapor. This is cause of the commonly observed "rocketing" of LP-Gas tanks which are suddenly ruptured. However, the remaining liquid vaporizes via heat transfer from its surroundings as does LNG. The vapor

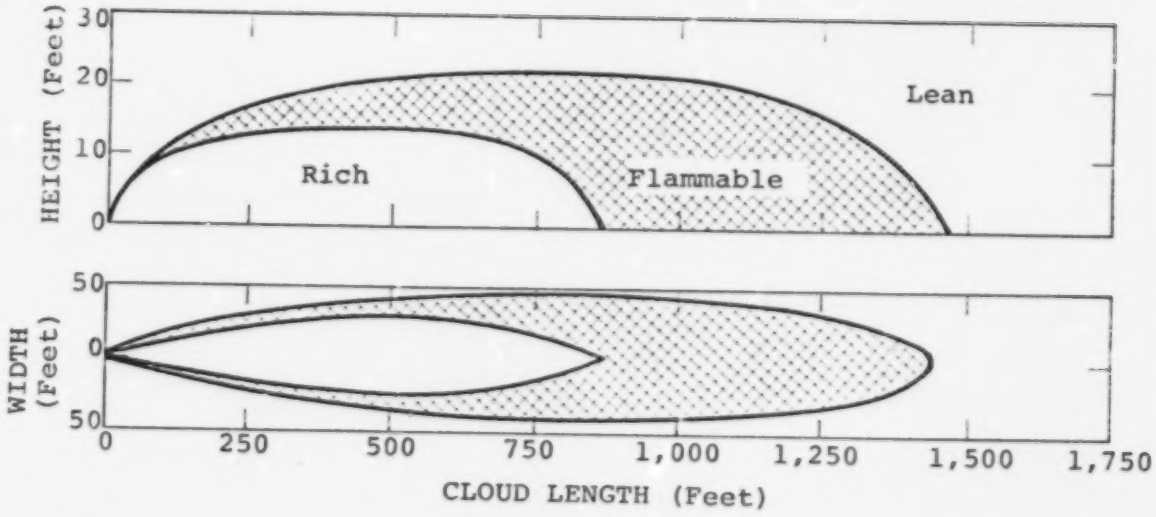


FIGURE 2. CONCENTRATIONS FROM A STEADY SOURCE

disperses in a negatively buoyant plume in much the same manner as does LNG vapor. Also similar to LNG vapor, the flammable portion of the LP-Gas vapor-air plume may be ignited readily, burning immediately. This in turn ignites any remaining liquid in a "pool fire."

## 3.2 Plume Detonation

If ignition is delayed, sufficient confinement may exist to cause initiation of detonation. In a well documented accident in a rural area near Port Hudson, Mo., investigation established that following a pipeline rupture a propane-air cloud covering several acres formed and detonated.[12] Concurrent with the rise of fireball, winds approaching 100 mph along the ground were generated. Damage corresponding to overpressures of 5 and 0.5 psi were observed at distances of 1,000 and 3,700 ft from the center of the blast. Fortunately no one was killed since the 24 minute time delay between the pipeline break and detonation permitted the few residents in the area to escape. However, their residences and farm buildings were severely damaged.

In other accidents, occurrence of detonation could not be fully proved, but overpressure effects have caused widespread property damage and many injuries. Such accidents include the explosion of vapor air clouds following tank car punctures in the rail yards at East St. Louis[13] and at Decator, Ill.[14] The number of fatalities caused by overpressure effects is not known but is believed to be relatively small.

Only the portion of the cloud that is within the flammable composition will detonate, generating overpressures of 250

to 300 psi. Because of the high overpressures, one may assume that all persons are killed and all ordinary buildings are destroyed within and somewhat outside the detonating portion of the cloud. The fuel rich portion of the cloud and the remaining pool of liquid then ignites and burns. Flammable property within and near this fire will be destroyed as indicated below. These consequences correspond to the top box in the right side of Figure 1.

### 3.3 Flash Fire

It detonation does not occur, the plume simply deflagrates (flash fire) and may ignite secondary fires. Any residual liquid at the spill site also ignites, and additional fires in surrounding structures may be ignited via flame impingement and thermal radiation. Buildings within the plume may be destroyed by explosion if prior to ignition they were infiltrated by a flammable mixture. Accounts of these flash fires indicate persons within the flammable plume will be killed whereas people outside will receive at most, serious burns. These consequences correspond to the second box down from the top at the right side of Figure 1. In contrast to detonation, the injury and damage mechanisms are burns and fires; overpressure effects are less important.

This accident chain is a major cause of fatalities from LP-Gas leaks and spills. For 50 accidents occurring during the years 1931 to 1961 (cf. Table XI in Appendix B) 60 persons (non-employees) were killed by flash fires in LP-Gas vapor plumes, which is 76 percent of the non-employee fatalities in the 50 accidents. Notable examples of this type of accident occured at Berlin, N.Y. in 1962[15] and near Meldrim, Ga. in 1959.[16] The former accident

involved a 7,000 gal tank truck which hit a tree near a curve, rupturing the tank. All of the LP-Gas vaporized, forming a plume which covered 5 acres. The subsequent flash fire and explosions in houses and other buildings killed ten persons. The second accident involved the derailment and subsequent puncturing of a LP-Gas rail tank car. The vapor-air plume spread over part of an adjacent picnic ground before igniting. Twenty-five persons were killed.

Another example was the fire following the ramming of a British tankship, the MV ALVA CAPE, by the tankship SS TEXACO MASSACHUSETTS in New York harbor, June 16, 1966. A cargo of naptha spilled onto the water between the two ships and survivors remembered seeing a large vapor cloud forming. Approximately two minutes after the collision, an explosion on a tug between the two ships ignited the vapor and the "pool". Thirty-three persons, all crew members of the ships and tugs involved, died, some in the initial flash fire and explosions, some trapped in the burning "pool" fire in the water (especially the crew of one of the tugs) and others in the fires on board the ALVA CAPE and the TEXACO MASSACHUSETTS. All damage and injuries were confined to the ships and their crews. Because of the similarities of naptha and LNG vapors, the events of this accident are suggestive of the events that could occur as the result of a ramming of a LNG tankship.

### 3.4 Fixed and Spreading Pool Fires

The third and fourth boxes on the right in Figure 1 correspond to what are expected to be the more common events: immediate ignition followed by a "pool" fire, if the spill involved the

rupture of a large container, or a torch fire, if piping is broken. A large LNG spill on water could create a "pool" fire a few thousand feet in diameter because of the rapid spread of LNG on water. Thus, the destruction of flammable harbor and industrial facilities by fire within this range is expected. Fatalities are not generally expected unless the victims somehow become entrapped in or near the pool, as might be the case of an LNG tanker crew (cf. the above example).

The most notable example of a spreading "pool" fire (and coincidently the only known major catastrophe in the U.S. involving LNG) occurred in Cleveland in October, 1944. A storage tank containing 38,000 bbls of LNG collapsed. The escaping liquid caught fire soon afterwards, and as it burned, it flowed down nearby streets, across parking lots, and into basements and storm sewers.[18] Accumulations of vapor in the latter regions caused serious explosions which demolished many homes and buildings. Fire also spread by thermal radiation from the flame of the burning liquid pool near the collapsed tank. The flame oscillated in "bursts" and reached a height of 2800 feet. Combustibles were ignited at distances of more than 1000 feet away (by radiation). In all, 135 people were killed and approximately $10,000,000 property damage was incurred. This accident was attributed to a material failure (brittle fracture) of the tank wall at the cryogenic temperature. An important contributing factor was that the tanks were not diked to prevent the spread of the flammable liquid.

The likelihood of this type of accident has been substantially reduced by the design of modern LNG tanks. The ability of materials now used for tank construction to serve at cryogenic

temperatures has been thoroughly demonstrated. Also, storage tanks will be diked to prevent the spread of LNG in the event the tank does rupture. However, the use of diking procedures for LNG tankers while in port apparently are not planned.

### 3.5 Exploding Tanks

The two boxes at the bottom on the right hand side of Figure 1 correspond to fatalities caused by exploding tanks and other pressure vessels in a fire. For the 50 LP-Gas accidents during the period 1931 to 1961, 24 percent of the non-employee fatalities (19 persons) were caused by this mechanism. The tank, whose wall has been overheated and weakened by an impinging flame (often caused by an LP-Gas torch from broken piping), ruptures. The high pressure propells the liquid in all directions along with many high-velocity fragments from the tank. A large fireball forms immediately as the liquid flashes to vapor and is ignited by the flame. The victims often are firemen who are trying to control or extinguish the initial fire. Two notable examples of this occurred at Deer Lake, Pa., in 1959[19] and at Kingman, Arz., in 1973.[20] In the former accident a collision with an auto broke some piping on a 7,000 gal LP-Gas tank truck. The escaping gas ignited and the flame impinged on the upper portion of the rear tank head. Approximately 45-minutes later the tank ruptured and rocketed forward 900 feet, killing 11 spectators. At Kingman, Arz., an unloading line connection failed and the torch of burning gas impinged on a 34,000 gal LP-Gas rail tank car. The tank car ruptured 19 minutes later, creating a 200-foot diameter fireball which killed 12 firemen.

## 3.6 Flammable Plumes – The Major Risk of Fatalities

Based on the accident experience discussed above, it appears that the flash burning of the vapor plume is the most important mechanism causing fatalities among the general population from LP-Gas accidents. This is the reason for treating only this mechanism in this analysis of risk. The event sequence associated with this mechanism is that having the events "Vapor-Air Plume Forms" and "Delayed Ignition" in Figure 1.

This also is expected to be the most important fatality mechanism for LNG fires. The other mechanisms, although they may contribute to risk in some situations, are presently thought to be of lesser importance for the following reasons.

Prerequisites of the exploding tank mechanism are either the flammable liquid is stored in a pressure vessel or the tank is heated so rapidly that upon rupture much of the liquid flashes to vapor and is thrown out in all directions. LNG is not stored in pressure vessels and situations in which rapid super-heating occurs are not easily visualized since the vessels are insulated. Thus, the exploding tank mechanism generally is not expected for LNG. With regard to the overpressure mechanism, air-methane mixtures are known to be considerably more difficult to detonate than mixtures of air and propane or other higher hydrocarbons. Consequently, high-overpressure effects presently are not believed to be a major fatality-producing mechanism in LNG spills. However, this statement is tentative since there is some fragmentary evidence that the deflagration of very large flammable clouds may create significant overpressures.[2] With regard to pool

fires, the spreading pool fire of the type which occurred at Cleveland is believed to be the more dangerous. It is thought that whereas some people might become entrapped in a rapidly spreading and burning liquid, they could evade much of the thermal radiation from a fixed pool fire. However, the use of dikes and catch basins about LNG tanks and piping should reduce the possibility of spreading pool fires at LNG terminals. Spreading pool fires from tankship ruptures on water are expected to endanger primarily the tankship's crew and not the general population on land.

## 4.0 Risk Assessment Methodology

The method for estimating the probability and number of fatalities in the general population caused by a flash fire in a vapor-air plume is described in this section. The number of fatalities is assumed to be the product of (1) the ground area, A, covered by the flammable portion of the plume at the moment of ignition and (2) the population density, $\rho$, at the site of the spill. As will be described, A is related to the vaporization rate, Q, a parameter characterizing the dispersion of the vapor, V, and a parameter, I, relating flammable plume area and ignition.

The relationship between the number of fatalities and the above parameters was modeled in effect as an event tree. This tree is shown in Figure 3 where for convenience the four sets of branch points corresponding to the above four parameters* have been condensed into blocks. The course of an accident event sequence, beginning with the spill and ending with fatalities is determined by the value of each parameter prevailing at the time. The frequency of occurrence of that accident event sequence is the joint frequency of the particular parameter values: $P(Q) \cdot P(V) \cdot P(I) \cdot P(\rho)$. Generally, a given number of fatalities results from more than one accident event sequence, and consequently, the frequency of that number of fatalities is the sum of the frequencies of the individual event sequences.

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*In order to have a finite number of branch points, continuous parameters were discretized into a finite number of ranges (e.g., two values for Q) as will be described.

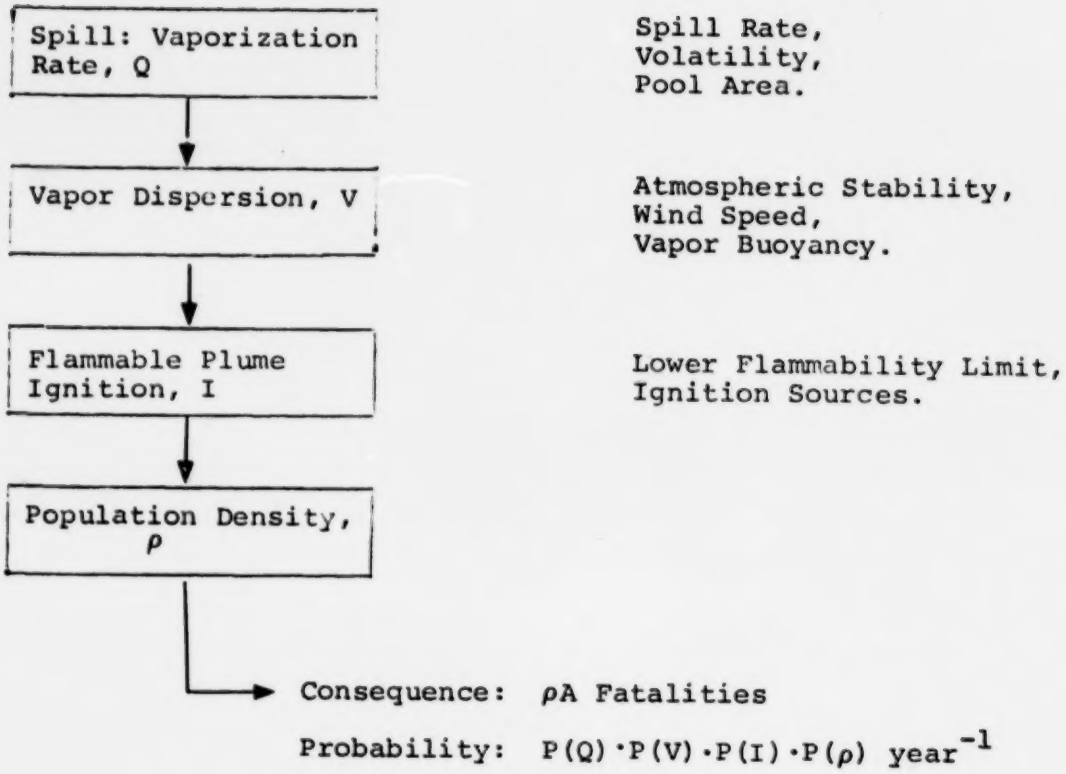
Spill: Vaporization Rate, Q
Spill Rate, Volatility, Pool Area.
Vapor Dispersion, V
Atmospheric Stability, Wind Speed, Vapor Buoyancy.
Flammable Plume Ignition, I
Lower Flammability Limit, Ignition Sources.
Population Density, ρ
Consequence: ρA Fatalities
Probability: P(Q)·P(V)·P(I)·P(ρ) year$^{-1}$

FIGURE 3

ACCIDENT-CONSEQUENCE BLOCK DIAGRAM

(CONDENSED EVENT TREE)

## 4.1 Spills, Vaporization Rate

For LP-Gas truck transport, typical accidental spills and their frequency of occurrence were garnered from accident records compiled by the Department of Transportation for the period July 1971 through December 1973.[21] The basic data for spills resulting from vehicular and loading-unloading accidents are collected in Table III (presented in Appendix A). Examination of these data suggests that the spills may be divided roughly into two categories: a steady leak and an instantaneous spill of the entire cargo. The former is assumed to be 100 gpm, which would be the nominal flow from a broken 2 or 3-inch unloading line. The size of the latter is assumed to be 4,370 gal, the average tank truck size.[22] Considering only losses of 100 gals or more, the frequency of the leak-type is 7.6 year$^{-1}$ (19/2.5); the frequency of the instantaneous spill is 2.8 year$^{-1}$ (7/2.5). However, the data in Table III does not include all significant accidents occurring in this 2.5 year period. It has been estimated that the reported accidents are only 20 percent of the actual number.[23] Therefore the above frequencies were multiplied by a factor of five to give P(Q) for tank trucks:

| | |
|---|---|
| 100 gpm leak accidents, | 38 year$^{-1}$, |
| 4,370 gal "instantaneous" spills, | 14 year$^{-1}$. |

To obtain what is believed to be an upper bound, the frequency of accidental spills from LNG tankships was assumed to be the same as for present day petroleum tankships. Accident data for the latter were obtained from records kept by the U.S. Coast Guard.[24] Table IV lists all collisions and groundings of petroleum tankships which occurred in U.S.

TABLE IV

TANKSHIP INCIDENTS IN U.S. HARBORS INVOLVING LOSS OF LIQUID CARGO
(1966 - 1972)

| Ship, Tonnage | Date | Damage and Cause | Amount Lost | Time Period of Loss |
|---|---|---|---|---|
| A. D. DUMONT 500-1,000 tons | 10/14/72 Day | Cracks in #3 and #4 tanks, grounding | ~100 bbls | Unknown |
| IDAHO STANDARD 10,000-15,000 tons | 2/1/72 Night | Slight tears in #2 and #7 center tanks, grounding | 500 bbls | 8 hours |
| ESSO GETTYSBURG over 15,000 tons | 1/23/73 Night | Grounding | 386,000 gal. | Unknown |
| OREGON STANDARD 10,448 tons | 1/25/71 Night (Fog) | Collision; #2, #3, #4 port tanks ruptured | 800,000 gal. | <7 hours |
| WASHINGTON STANDARD 10,000-15,000 tons | 2/5/70 Day | Fore tanks holed by ice; 6'8" x 24" hole | sea water ballast | Unknown |
| EXXON MIAMI <15,000 tons | 3/9/71 | #8 starboard wing tanks leaked at seams; collision with dock | <50 bbls | Unknown |
| MARY A. WHALER 500-1000 tons | 12/23/68 Night | Damage to bottom #6 and #4 tanks | 100 gal. | 79 hours |
| TEXACO NEBRASKA 10,000-15,000 tons | 3/21/67 Day | Fracture in plating of bunker tank; grounding | minor leak | Unknown |
| ANGELO-PETRI 5,000-10,000 tons | 2/1/68 Night | Collision with barge; damage 6' inboard | 1,000 gal. | 3 hours |
| TEXACO FLORIDA 10,000-15,000 tons | 4/24/68 day | Hull fracture #1, #2 center tanks, 20' x 8" and 12' x 2' | $15,000 refined pet products. est. 3,000 bbls. | Unknown |

TABLE IV (continued)

| Ship, Tonnage | Date | Damage | Amount Lost | Time Period of Loss |
|---|---|---|---|---|
| ALVA CAPE<br>11,252 tons | 6/16/66<br>Day | No. 1 tank holed in collision | 8,000 to 9,000 bbls of naptha | A few minutes |
| F. A. VERDON<br>750 tons | 6/22/71<br>Day | Not available | 1,500 bbls. (est) | Not available |

ports during the period June 1966 through June 1972 and in which liquid cargo was lost. Based on these data it was assumed that LNG tankship losses would be of two types: a leak of 1,000 bbls/hr or the instantaneous rupture of one 25,000 $m^3$ tank, one of the five spherical tanks of the 125,000 $m^3$ (total) General Dynamics LNG tankship. Considering only tankships greater than 7,500 tons, the data in Table IV indicates that there are an average of 1 leak-type loss per year (6 losses/6 years) and 0.5 instantaneous losses per year (3 losses/6 years). For the latter, the accidents involving the ESSO GETTYSBURG, OREGON STANDARD AND ALVA CAPE were considered to be instantaneous.

There is, of course, the possibility that more than one of the five LNG tanks could be ruptured in an accident and contribute to the spill. This was not considered here. For the General Dynamics tankship, that more than one tank could be damaged by collision must have a lower probability since the tanks are spherical (not contiguous as are the rectangular tanks of a conventional tankship) and are arranged in a fore-and-aft line. On the other hand, a second tank could be caused to rupture via a "pool fire" following an accident and rupture of the first tank, but the existence of the fire precludes formation of a flammable plume.

For the years 1971 and 1972, tankships $\geq$ 7,500 tons made 30,095 and 35,562 trips into U.S. ports.[25] Thus we calculate that the average loss rate is $3 \times 10^{-5}$ per trip for leaks and $2 \times 10^{-5}$ per trip for instantaneous tank ruptures. As mentioned in Section 2.0, the expected number of LNG tankship trips into U.S. ports during 1980-

1985 is 999. Rounding this off to 1,000 trips per year, we obtain for the average annual loss rates P(Q) for LNG tankships:

| | |
|---|---|
| 1,000 bbls/hr leak accidents, | $0.03\ \text{year}^{-1}$, |
| 25,000 $m^3$ instantaneous spills, | $0.02\ \text{year}^{-1}$. |

The modeling of these spills as sources of vapor was as follows. The LP-Gas spills all were considered as point vapor sources. For the steady leak, the rate of vapor generation was assumed to be the same as the liquid leak rate, 100 gpm = 3,526 g/sec. For the instantaneous spill, an initial temperature of 68°F was assumed and accordingly 31 percent of the spill, 1,350 gal = $2.6 \times 10^6$ g, flashes into vapor. The remaining liquid which vaporizes at a more-or-less steady rate was ignored since conditions generally are such that it forms a smaller flammable plume than does the initial "puff."

The LNG spills, since they would occur on water, were modeled as area sources. The size of the spill area was estimated using the average of the data in References (8) and (10) for the LNG evaporation rate, 0.0385 $\text{lbs/ft}^2\text{-sec}$. For the steady leak, the vaporization rate was the same as the leak rate, 1,000 bbls/hr = $1.87 \times 10^4$ g/sec. The vaporization area was calculated to be circular with a diameter of 5.4 m.

For the instantaneous spill, the spill area and hence the rate of vaporization were assumed to increase, as $t^2$. Vaporization was assumed to be complete at the time the maximum area is attained, t = 726 sec and A = $2.4 \times 10^5$ $m^2$

(calculated from the relations in Reference 8). From these assumptions the vaporization rate 87.09 $t^2$ is derived, where t is the time from the beginning of the spill. The area assumed for the entire vaporization period was the average value for which the diameter is 392 m. The assumptions employed are approximations. For example, subsequent to the time of maximum pool diameter, some vaporization must occur but at a rapidly decreasing rate since the pool is observed to break up into rapidly diminishing smaller pools. However these approximations are not critical and as will be shown below the calculated flammable plume area is insensitive to them.

## 4.2 Vapor Dispersion

Atmospheric dispersion of the vapor was modeled with the appropriate Gaussian plume formulas as will be described in the following section. Experiments by the Bureau of Mines[9] and Esso Research and Engineering[10] indicated that the Gaussian formulas can be used for these applications with some success.

According to the Gaussian model, the parameters controlling dispersion are wind speed and atmospheric stability. For the latter, the Pasquill categories were used as a matter of convenience. The probability of a given dispersion characteristic, P(V), therefore is given by the joint probability distribution for the occurrence of wind speed and Pasquill stability category. These were obtained from U.S. Weather Bureau Data. For LP-Gas tank truck spills, the probabilities used for chlorine spills in the Eastern U.S. were assumed to be applicable here and are given in Table V. The joint probability distribution for the LNG tankship spills is similar and is given in Table VI. This distribution is the average of the data from the Newark and New Orleans Airports.

TABLE V. Joint Probability Distribution of Wind Speed and Pasquill Stability Categories (Used for the LP-Gas Spills)

| Wind Speed Range (kts) | Probability of Occurrence | | |
|---|---|---|---|
| | A-B-C | D | E-F |
| 0-3 | 0.0211 | 0.0163 | 0.0780 |
| 4-6 | 0.0504 | 0.0794 | 0.1495 |
| 7-10 | 0.0768 | 0.1960 | 0.0727 |
| 11-16 | 0.0121 | 0.2070 | - |
| 17-21 | 0.0006 | 0.0326 | - |
| Total | 0.1610 | 0.5313 | 0.3002 |

TABLE VI. Joint Probability Distribution of Wind Speed and Pasquill Stability Categories (Used for LNG Tankship Spills)

| Wind Speed Range (kts) | Probability of Occurrence | | |
|---|---|---|---|
| | A-B-C | D | E-F |
| 0-3 | 0.0172 | 0.0129 | 0.0707 |
| 4-6 | 0.0543 | 0.0744 | 0.1543 |
| 7-10 | 0.0806 | 0.1848 | 0.0834 |
| 11-16 | 0.0135 | 0.2091 | - |
| 17-21 | 0.0006 | 0.0382 | - |
| Total | 0.1662 | 0.5194 | 0.3084 |

The Pasquill formulas for the horizontal dispersion coefficient used in this analysis are those given in graphical form by Turner[26] (x is the downwind distance in meters):

TABLE VII. DISPERSION COEFFICIENTS

| Stability Class | $\sigma_y$ (meters) |
|---|---|
| A-B-C | $0.20 \times^{0.91}$ |
| D | $0.13 \times^{0.91}$ |
| E-F | $0.077 \times^{0.91}$ |

Here classes E and F have been averaged since much of the data reported does not distinguish between these classes. Also classes A, B and C were lumped together as class C for convenience. These classes as a group do not characterize more than 10 to 20 per cent of the stability conditions and of this fraction, class C predominates.

Both the Bureau of Mines[9] and the Esso Research and Engineering[10] experiments showed that the LNG vapor plumes were negatively buoyant. Following Reference (9), to account for this we have assumed

$$\sigma_z = 0.2\sigma_y \qquad (1)$$

for all stability classes.

## 4.3 Flammable Plume Area and Ignition

By definition the ground area covered by the flammable plume is that area enclosed by the isopleth corresponding to the lower flammable limit of the vapor at the instant of ignition.

For LP-Gas the lower flammable limit is 2.1 volume per cent (propane) and for LNG it is 5.0 volume per cent (methane). The maximum size of this area is related to the spill vaporization rate and the dispersion via a Gaussian plume formula. However, as discussed in Section 3.0, the probability that this area or some smaller area is attained at the time of ignition depends on the probability of ignition, which is P(I) in Figure 3. The treatment of these two subjects are discussed in the following paragraphs.

4.3.1 Maximum Flammable Area

Mathematically the ground area, A, is given by

$$A = \int_{x_{min}}^{x_{max}} 2y(x)\,dx \ , \qquad (2)$$

where x is the distance along the plume axis, y(x) is the lateral distance from the axis to the lower flammable limit isopleth, and $x_{min}$ and $x_{max}$ are the minimum and maximum distances at which the isopleth crosses the x-axis. The function y(x) is given by a Gaussian plume formula.

For the LNG tanker spills, it was assumed that the tanker was at least 100 m from shore (i.e. in the middle of a channel 200 m wide. Thus $x_{min}$ = 100 m for the steady leak and $x_{min} \geq 100$ m for the instantaneous spill. (Once vaporization is complete, and the puff drifts downwind, the distance from the spill to its trailing edge eventually exceeds 100 m.)

For LP-Gas leaks, the well known, point source steady plume formula for ground level concentrations was used:

$$\chi = \frac{Q}{\pi \sigma_y \sigma_z U} \exp\left[-\frac{1}{2}\left(\frac{y}{\sigma_y}\right)^2\right] . \qquad (3)$$

Here Q is the vapor source strength (3526 g/sec), U is the wind velocity (m/sec), and $\chi$ is the lower flammable limit (41.1g/$m^3$). The units of $\sigma$ and y (defined above) are meters. Flammable plume area computed as a function of wind speed and stability category are listed in Table VIII.

For the instantaneous LP-Gas spill the following standard equation was used for ground level concentrations:

$$\chi = \frac{2Q}{(2\pi)^{3/2} \sigma_y \sigma_x \sigma_z} \exp\left\{-\frac{1}{2}\left(\frac{x-ut}{\sigma_x}\right)^2 - \frac{1}{2}\left(\frac{y}{\sigma_y}\right)^2\right\} . \qquad (4)$$

Here Q is the total weight that flashes into vapor, 0.31 x 4370 gal. = 2.59 x $10^6$g, and $\chi$ has the same value as above, 41.1 g/$m^3$. This equation has circular isopleths centered at x = ut. Assuming that $\sigma_x = \sigma_y$ and recalling that $\sigma_z = 0.2\sigma_y$, the area enclosed by an isopleth is

$$A = \pi \sigma_y^2 \; 2 \ln\left(\frac{10Q}{\chi \sigma_y^3 (2\pi)^{3/2}}\right) . \qquad (5)$$

As will be explained in the following subsection on ignition probability, we are interested primarily only in the maximum area of the vapor puff. This is easily computed from Equation (5):

$$A_{max} = \frac{3}{2}\left(\frac{10Q}{\chi}\right)^{2/3} e^{-1} . \qquad (6)$$

TABLE VIII

LP-Gas Flammable Plume Area, 100 GPM Spill Rate

| Wind Speed[a] (m/sec) | Plume Area ($m^2$) | | |
|---|---|---|---|
| | A-B-C | D | E-F |
| 0.76 | 1750 | 2879 | 4917 |
| 2.54 | 495 | 814 | 1391 |
| 4.32 | 284 | 467 | 797 |
| 6.86 | 175 | 287 | b |
| 9.65 | 122 | 201 | b |

a These are the mean values of the wind speed ranges in Tables V and VI.

b These conditions do not occur.

It is noteworthy that $A_{max}$ is independent of both stability category and wind speed. This is because the diffusion is independent of wind speed and the stability categories determine only the downwind location at which the maximum occurs. Substituting the values for $\chi$ and Q, we find that $A_{max} = 4.06 \times 10^3 m^2$ (approximately one acre).

Equation (3) was used for the LNG tank leaks with $Q = 1.87 \times 10^4 g/sec$ and $\chi = 33.9\ g/m^3$. As mentioned above in Section 4.1 an area vaporization source was assumed which was treated by a method suggested by Turner.[26] An upwind virtual source is assumed for the vaporizing pool such that the value of $\sigma_y$ at the center of the pool equals the pool radius. Thus to evaluate $\sigma_y$ by the formulas in Table VII the virtual distance, $x_o$, must be added to the downwind distance. Since the pool diameter was estimated to be 5.4 m,

$$x_o = \left(\frac{5.4}{2Y}\right)^{1/0.91} , \qquad (7a)$$

and

$$\sigma_y = Y\,(x + x_o)^{0.91} , \qquad (7b)$$

where Y is the appropriate coefficient from Table VII. The computed flammable plume areas are listed in Table IX.

Since the instantaneous LNG spills vaporize over a considerable period of time, the spreading disk model of a Gaussian plume was used.[27]

TABLE IX

LNG Flammable Plume Area from a Steady 1000 bbl/hr. Spill Rate (Area within 100m of the spill is excluded).

| Wind Speed[a] (m/sec) | Flammable Plume Area ($m^2$) | | |
|---|---|---|---|
| | A-B-C | D | E-F |
| 0.76 | 6941 | 14943 | 28934 |
| 2.54 | 72 | 1757 | 5245 |
| 4.32 | 0 | 154 | 1710 |
| 6.86 | 0 | 0 | b |
| 9.65 | 0 | 0 | b |

a These are the mean values of the wind speed ranges in Tables V and VI.

b These conditions do not occur.

The continuous series of instantaneous puffs (Equation (4)) was integrated over the time of vaporization:

$$\chi = \int_0^{t_m} \frac{2(87.09t'^2)dt'}{(2\pi)^{3/2}\,\sigma_y\sigma_z\sigma_x} \exp\left\{-\frac{1}{2}\left(\frac{x-u(t-t')}{\sigma_x}\right)^2 - \frac{1}{2}\left(\frac{y}{\sigma_y}\right)^2\right\} \quad (8)$$

As mentioned in Section 4.1 $t_m$ is the time for complete vaporization, 727 sec, and $87.09t'^2$ g/sec is the instantaneous vaporization rate. Performing the integration with $\sigma_x = \sigma_y$ and $\sigma_z = 0.2\sigma_y$, the complicated expression, Equation (9) was obtained:

$$\chi = \frac{219 \exp\left[-\frac{1}{2}\left(\frac{y}{\sigma_y}\right)^2\right]}{\pi\sigma_y^{\,2}\,U}$$

$$\left\{\left[\left(\frac{x}{u}-t\right)^2 + \left(\frac{\sigma_y}{U}\right)^2\right]\left[\operatorname{erf}\left(\frac{x-u(t-t_m)}{\sqrt{2}\,\sigma_y}\right) - \operatorname{erf}\left(\frac{x-ut}{\sqrt{2}\sigma_y}\right)\right]\right.$$

$$+\sqrt{\frac{2}{\pi}}\,\frac{\sigma_y}{u}\left(\frac{x}{u} - (t+t_m)\right)\exp\left[-\frac{1}{2}\left(\frac{x-u(t-t_m)}{\sigma_y}\right)^2\right]$$

$$\left.-\sqrt{\frac{2}{\pi}}\,\frac{\sigma_y}{u}\left(\frac{x}{u}-t\right)\exp\left[-\frac{1}{2}\left(\frac{x-ut}{\sigma_y}\right)^2\right]\right\}. \quad (9)$$

This equation in effect gives y as a function of x and t (time after the beginning of the spill). Since the spill is an area source with a diameter of 392 m, the virtual source distance upwind of the spill was

$$x_o = \left(\frac{392}{2Y}\right)^{1/0.91}, \qquad (10a)$$

and

$$\sigma_y = Y\,(x + x_o)^{0.91}\,. \qquad (10b)$$

Flammable plume areas were computed as a function of time with the aid of a computer (program listing given in Appendix C). Table X lists the maximum values as a function of wind speed and stability category. All of the values fall into the range of $1 \times 10^6$ to $1.6 \times 10^6$ $m^2$. This result is not too surprising in view of the above remarks about the maximum areas of instantaneous, point source puffs. It is interesting to note that treating the 25,000 $m^3$ LNG spill as an instantaneous, point source release, the maximum area by Equation (5) is $1.22 \times 10^6$ $m^2$, in good agreement with the values from Equation (9).

### 4.3.2 Reduced Areas - Ignition Probability

As explained in Section 3.0, it is rarely expected that a flammable plume will achieve its maximum size, since ignition is likely, especially in populated areas. The great variety of human activities generate a myriad ignition sources: open flames, stray electrostatic sparks, arcing of electrical switches and motor brushes, cigarettes, electrical and gas ranges, hot water heaters, electric starting circuits on automobiles, etc. These conspire to make the value of P(I) in Figure 3 much less than unity for large flammable plumes. Hence, the actual ground area involved in the flash burning of a plume generally is much less than the maximum area calculated above.

TABLE X

LNG Flammable Plume Areas From a 25,000$m^3$ Spill

(Any area within 100m of the spill is excluded.)

| Wind Speed[a] (m/sec) | Flammable Plume Area ($10^6$ sq. meters) | | |
|---|---|---|---|
| | A-B-C | D | E-F |
| 0.76 | 1.21 | >1.1[c] | 1.1[c] |
| 2.54 | 1.32 | 1.36 | 1.38 |
| 4.32 | 1.35 | 1.50 | 1.57 |
| 6.86 | 1.19 | 1.39 | b |
| 9.66 | 1.03 | 1.30 | b |

a These values of the average of the ranges in Tables V and VI.

b These conditions do not occur.

c Maximum values were not computed but are not much larger than the values given.

For an area comprising a given population density and type of activity (e.g., industrial, commercial, etc.), it is reasonable to suppose that the probability of ignition depends only on the area covered by the flammable portion of the plume. Accounts of many flash fires following spills of LP-Gas and other highly volatile fuels give data relevant to the flammable plume area at the time of ignition. Data from these accounts have been collected in Table XI (Appendix B) for 59 incidents. Besides several details about each incident, the estimated flammable plume area is listed in each case. In a few accounts the plume area was stated and is the value listed in the table. However in most cases, only the distance to the ignition source was indicated, in which case the area, A, was estimated using the formula

$$A = 0.175r^2, \tag{11}$$

where r is the stated distance. In effect the plume is assumed to be a segment of a circle with an included angle of 20° (corresponding approximately to the average of Pasquill D and C stabilities.

Fifty percent of the spills were ignited essentially immediately; plume area was less than 30 $m^2$. Only 7 percent of the spills had flammable plumes at ignition greater than 10,000 $m^2$ (2-1/2 acres), and none were larger than 30,000 $m^2$ (7-1/3 acres). This latter, upper limit probably resulted because of the maximum quantity spilled was a few tens of thousands of gallons. Figure 4 shows the data plotted on log-normal-probability coordinates. The data lie nearly on a straight line with the equation

$$P(A) = 1/2 \left\{1 + \text{erf}\left(\frac{\log A - 1.38021}{2.45318}\right)\right\} \tag{12}$$

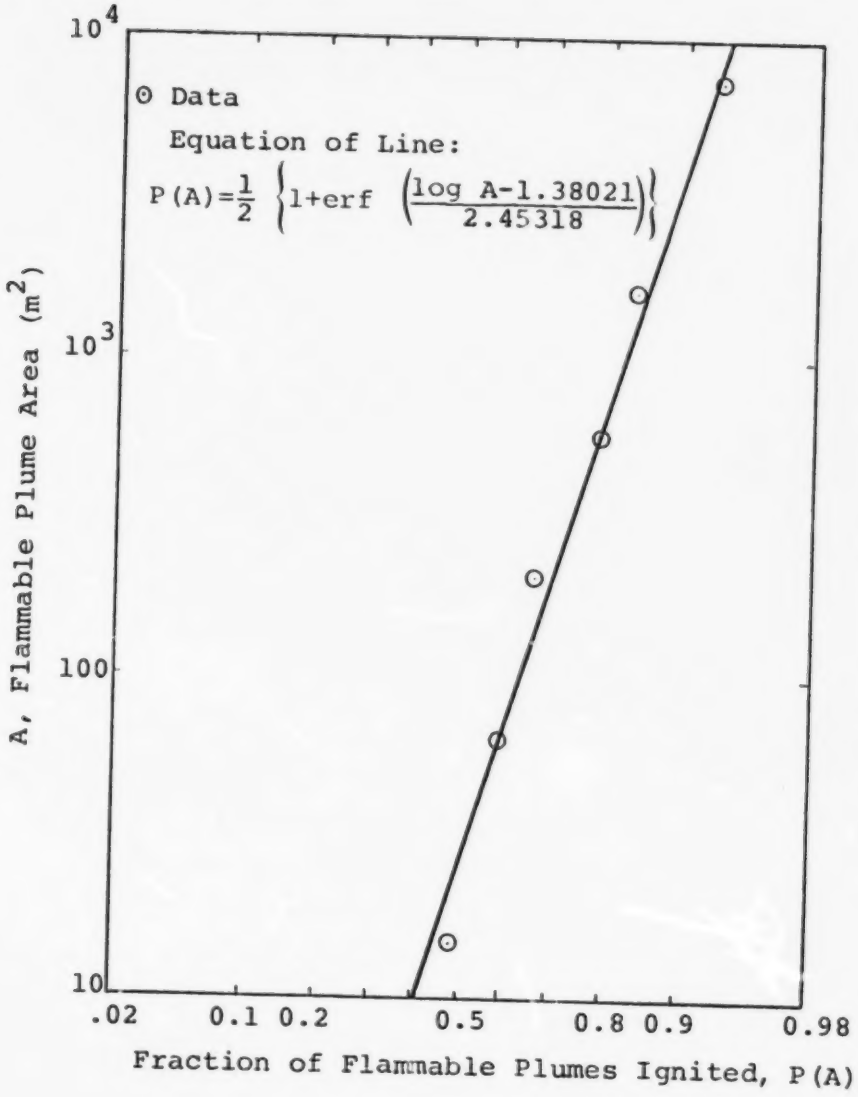


FIGURE 4

PROBABILITY OF FLAMMABLE PLUME IGNITION
VERSUS PLUME AREA AT THE TIME OF IGNITION

where P (A) is the cumulative probability that a plume of area A will have been ignited. Equation (12) predicts that the probability of a flammable plume being larger than 0.4 $mi^2$ (1 x $10^6 m^2$) is 0.004. The maximum flammable plume area attained from a 25,000 $m^3$ LNG spill on water is between 1.2 and 1.8 x $10^6$ $m^2$ (2/3 $mi^2$). Although flash fires in LNG vapor or methane-air plumes are not included in Table XI, Equation (12) is expected to apply to those plumes too because of the general similarity of the ignition properties of all flammable hydrocarbon-air mixtures.

In the calculation of risk it was convenient to divide up the possible plume areas into a set of ranges in which the maximum area of range was approximately 3 times the minimum area. The probability of ignition within that area range, the value of P(I) in Figure 3, is simply

$$P(I) = P(A_{max}) - P(A_{min}) . \qquad (13)$$

The area ranges and their corresponding ignition probabilities are listed in Table XII.

These values of P(I) were used in the following way. Suppose, for example, the maximum flammable plume area from a dispersion estimate was 28,934 $m^2$ (which is the area listed in Table IX for a LNG tankship leak under a 0-3 kt wind and E-F stability). The probability that the plume is ignited when it is in the range of 10,000 to 30,000 $m^2$ is only 0.0287. The probability that it is ignited before it reaches that size, say when it is <30 $m^2$ is 0.5223 and similarly for the in between size ranges. There is also a finite possibility that the plume will not be ignited at all: 1 - P(30,000) = 0.0367. Two examples of unignited plumes are mentioned in Table XI.

TABLE XII

Plume Area Ignition Probability

| Area Range ($m^2$) | Probability |
|---|---|
| <30 | 0.5223 |
| 30-100 | 0.1173 |
| 100-300 | 0.0969 |
| 300-1,000 | 0.0884 |
| 1,000-3,000 | 0.0615 |
| 3,000-10,000 | 0.0479 |
| 10,000-30,000 | 0.0287 |
| $30,000-1x10^5$ | 0.0183 |
| $1x10^5-3x10^5$ | 0.0094 |
| $3x10^5-1x10^6$ | 0.0052 |
| $1x10^6-3x10^6$ | 0.0022 |
| $3x10^6-1x10^7$ | 0.0011 |

In the case of the puff-type plumes from the more-or-less instantaneous LP-Gas and LNG spills, it is appropriate to use the area "swept out" by the flammable puff rather than the instantaneous area. This factor was ignored for the LNG spills, since the correction was negligible for all but the lowest wind velocities. For the LP-Gas "puffs" however, the correction was significant. Figure 5 shows the corresponding values of puff and swept areas used in the risk calculation. Swept area depends on atmospheric stability since this determines the distance downwind at which an area of a given size is attained.

## 4.4 Population Density

The population density distributions used in the estimations of risk are listed in Table XIII. The area fraction for a given population range corresponds to $P(\rho)$ in Figure 3. The distribution for highways and railroads is the same distribution used for chlorine spills in Reference 1. The distribution for U.S. ports is based on the population distribution surrounding Raritan Bay and the Arthur Kill in New York Harbor.

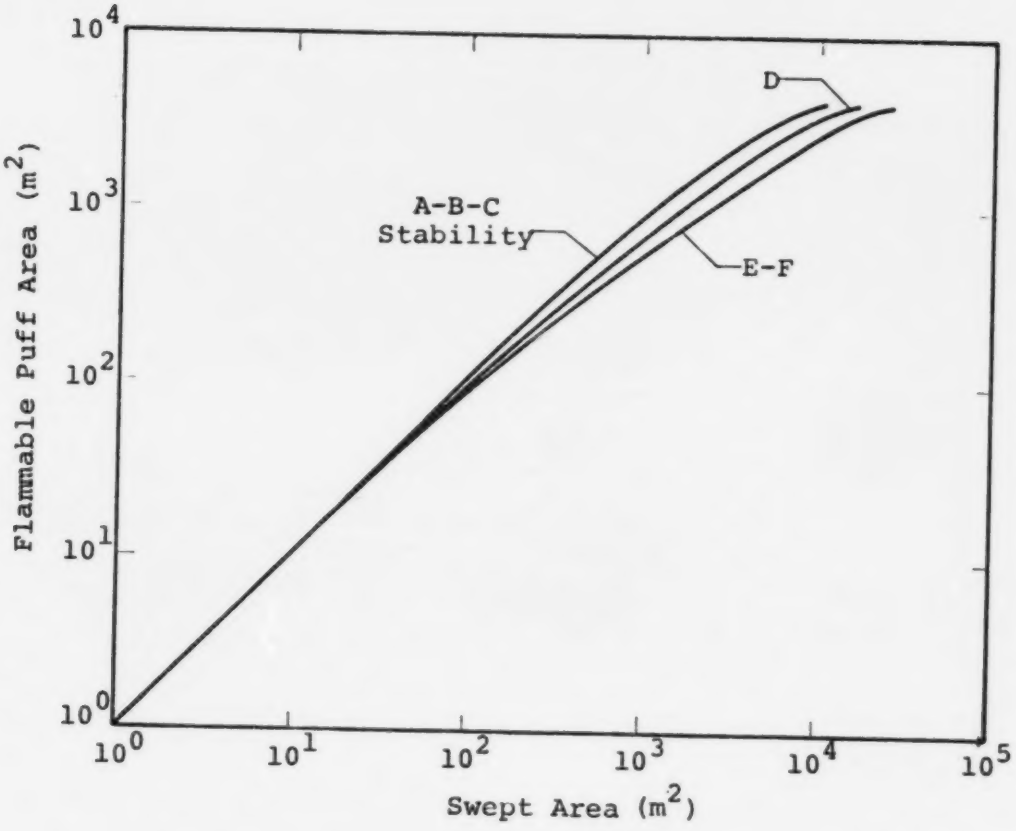


FIGURE 5

FLAMMABLE PUFF AREA VERSUS SWEPT AREA
FOR THE RUPTURE OF LP-GAS TANK TRUCKS

TABLE XIII

Population Density Distributions

| Highways and Railroads | | U. S. Ports | |
|---|---|---|---|
| Range Persons/mi$^2$ | Area Fraction, $P(\rho)$ | Range Persons/mi$^2$ | Area Fraction, $P(\rho)$ |
| <50 | 0.1894 | <10 | 0.0252 |
| 50-100 | 0.4796 | 10-30 | 0.0252 |
| 100-250 | 0.2246 | 30-100 | 0.0504 |
| 250-500 | 0.0277 | 100-300 | 0.0588 |
| 500-1000 | 0.0331 | 300-1000 | 0.1513 |
| 1000-2500 | 0.0200 | 1000-3000 | 0.2941 |
| 2500-5000 | 0.0171 | 3000-10,000 | 0.3782 |
| 5000-10,000 | 0.0083 | >10,000 | 0.0168 |
| >10,000 | 0.0003 | | |

## 5.0 Results

The consequent fatalities and their frequency were calculated for each of all the combinations of values of the variables (in Figure 3). The frequency of accidents with a number of fatalities within a threefold range were summed, yielding a histogram of accident frequency versus fatalities per accident. These are shown in Figures 6 and 7, LP-Gas tank truck accidents and LNG tankship accidents, respectively. Tables XIV and XV list the frequency of fatalities for the leak-type and instantaneous spills, separately.

Although LP-Gas accidents are more frequent (by a factor of 50 for the same fatality accident), the largest LNG accident has more fatalities, 10,000 to 30,000 versus 30 to 100. Of course this arises from the much larger LNG spills; maximum flammable plume size and maximum population density combine to yield maximum fatalities. In this connection it may be noted that leaking LNG tankships contributed negligibly to all but the lowest (<30) fatality accidents.

In the histogram form, the calculated results are difficult to compare with historical accident data. Such data often are statistically too few to resolve into an accident frequency versus fatalities relationship. However, the average annual fatalities from given types of accidents can be derived from accident data. Such a number is related to an integral of the histograms. As suggested by the manner in which they were calculated, the ordinate of the bars in the histograms are equivalent to the integral,

$$\int_{N_i}^{N_{i+1}} f(N)\,dN \quad ,$$

Table XIV

Predicted Frequency of Fatal
LP-Gas Tank Truck Accidents in the U.S.
(Based on Amounts Presently Shipped
by Tank Truck, About $20 \times 10^9$ Gal)

| Fatalities Per Accident | Frequency (per year) | |
|---|---|---|
| | Leak | Tank Rupture |
| 0.001 - 0.003 | 3.5 | 1.3 |
| 0.003 - 0.01 | 4.6 | 1.7 |
| 0.01 - 0.03 | 2.9 | 1.1 |
| 0.03 - 0.1 | 1.4 | 0.86 |
| 0.1 - 0.3 | 0.5 | 0.54 |
| 0.3 - 1.0 | 0.15 | 0.17 |
| 1.0 - 3.0 | 0.046 | 0.047 |
| 3.0 - 10 | 0.011 | 0.021 |
| 10 - 30 | $1.2 \times 10^{-3}$ | $3.3 \times 10^{-3}$ |
| 30 - 100 | $4.2 \times 10^{-5}$ | $1.2 \times 10^{-4}$ |

Table XV

Predicted Frequency of Fatal
LNG Tankship Accidents in U.S. Ports
(Based on 1,000 Tankship Trips to the U.S. in 1980)

| Fatalities Per Accident | Frequency (per year) | |
|---|---|---|
| | Leak | Tank Rupture |
| 0.01 - 0.03 | $3.5 \times 10^{-3}$ | - |
| 0.03 - 0.1 | $4.9 \times 10^{-3}$ | - |
| 0.1 - 0.3 | $1.8 \times 10^{-3}$ | $2.0 \times 10^{-3}$ |
| 0.3 - 1.0 | $1.2 \times 10^{-3}$ | $1.6 \times 10^{-3}$ |
| 1.0 - 3.0 | $7.3 \times 10^{-4}$ | $1.3 \times 10^{-3}$ |
| 3.0 - 10 | $4.3 \times 10^{-4}$ | $9.0 \times 10^{-4}$ |
| 10 - 30 | $1.7 \times 10^{-4}$ | $6.2 \times 10^{-4}$ |
| 30 - 100 | $3.3 \times 10^{-5}$ | $3.8 \times 10^{-4}$ |
| 100 - 300 | $1.2 \times 10^{-6}$ | $2.2 \times 10^{-4}$ |
| 300 - 1,000 | - | $1.1 \times 10^{-4}$ |
| 1,000 - 3,000 | - | $5.6 \times 10^{-5}$ |
| $3,000 - 1 \times 10^{4}$ | - | $1.8 \times 10^{-5}$ |
| $1 \times 10^{4} - 3 \times 10^{4}$ | - | $7.4 \times 10^{-7}$ |

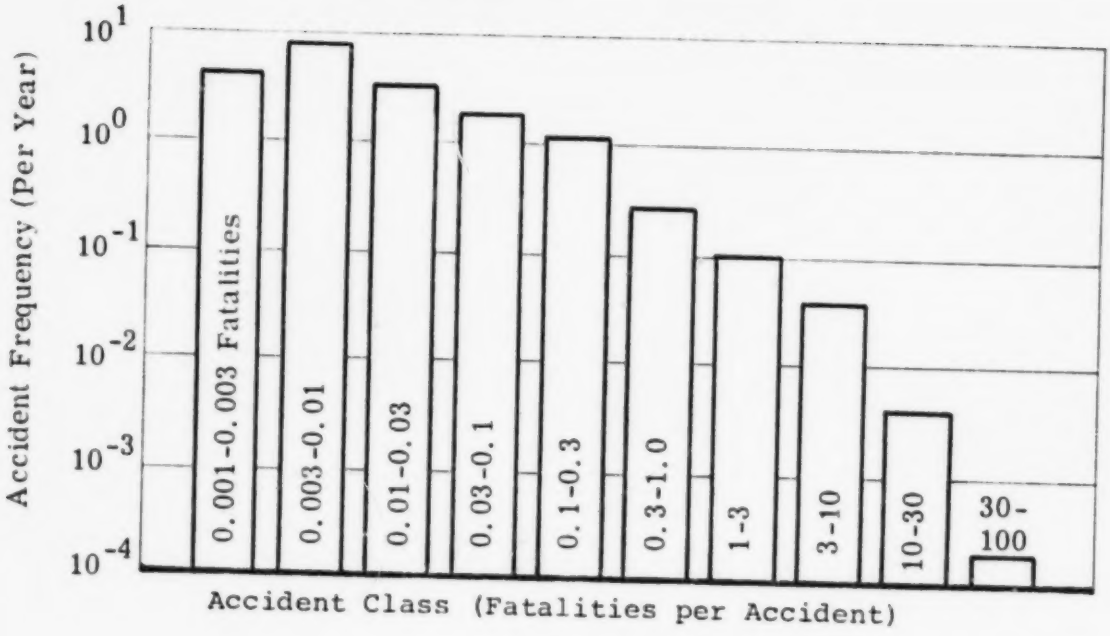


FIGURE 6: LP-GAS TANK TRUCK ACCIDENTS
(TANK PUNCTURES, UNLOADING ERRORS, ETC.)

PREDICTED FREQUENCY OF CLASSES OF ACCIDENTS
WITH A STATISTICAL NUMBER OF FATALITIES

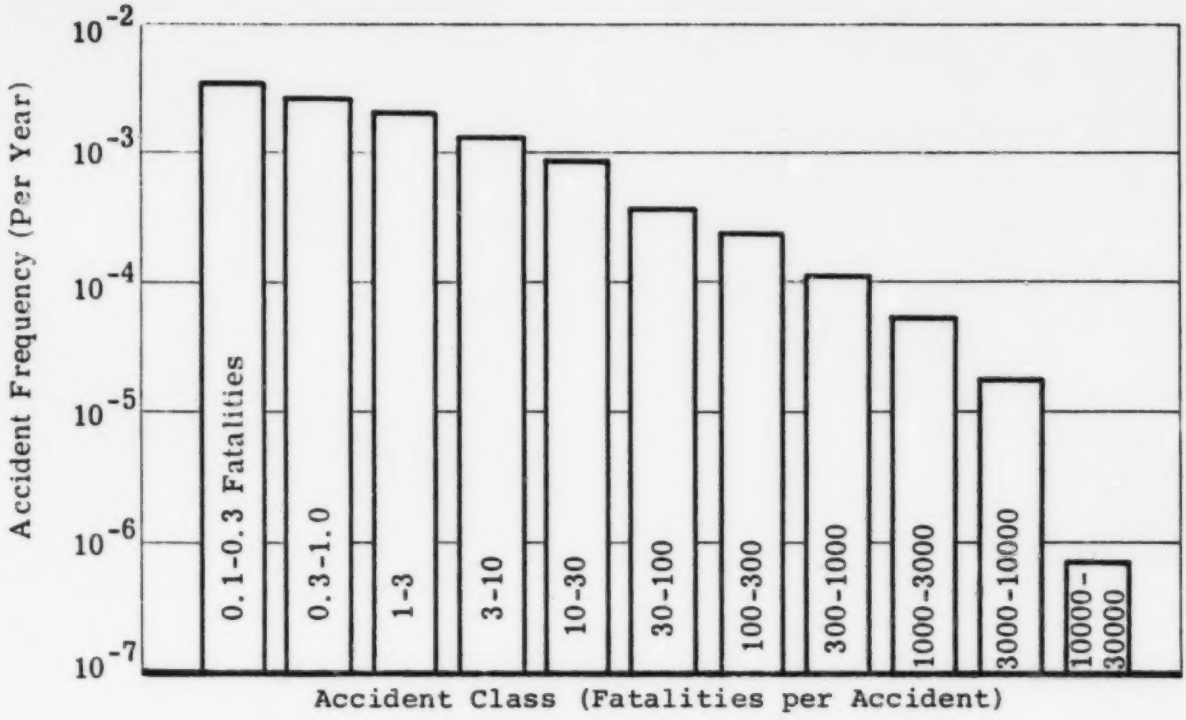


FIGURE 7: LNG TANKSHIP ACCIDENTS (COLLISIONS, GROUNDINGS, ETC.) ASSUMING 1,000 TANKSHIP TRIPS PER YEAR

PREDICTED FREQUENCY OF CLASSES OF ACCIDENTS WITH A STATISTICAL NUMBER OF FATALITIES.

when $N_{i+1}$ and $N_i$ are the maximum and minimum fatalities of the i th bar. The average annual fatalities, $N_t$, for accidents with fatalities, $\leq N_{i+1}$, then may be defined as

$$N_t = \sum_{N_i=N_{min}}^{N_{i+1}} <N> \int_{N_i}^{N_{i+1}} f(N)\,dN \quad , \tag{13}$$

where $<N> = (N_i+N_{i+1})/2$ is the average number of fatalities of the bar and $N_{min}$ is the minimum number of fatalities for all the accidents. Plots of $N_t$ versus $<N>$ are shown in Figure 8. The value of $N_t$ corresponding to summing to $N_{max}$, the fatalities of the maximum accident, is the average annual number of fatalities expected for that type of accident.

For LNG tankship accidents, the predicted statistical fatalities are 0.4 per year for 1,000 tanker trips, annually, to U.S. ports. Of course, this number is not easily compared with historical data. It is a projection for the 1980-1985 time period, and actual experience with LNG tankship activity in U.S. ports is not significant, statistically. It should be kept in mind that the LNG tankship number is believed to represent an upper bound since it is based on a number of conservative assumptions, the most significant of which is that no collision protection was assumed for the double hull structure used in LNG tankships (see Section 6.0).

There are historical fatality data for LP-Gas tank truck accidents. During the period 1931 to 1961, 37 persons (non-workers) were killed in LP-Gas flash fires and explosions from accidents involving tank trucks. This is an average of 1.23 fatalities per year (37 fatalities/30 years), which agrees remarkably well with the calculated value of 1.2.

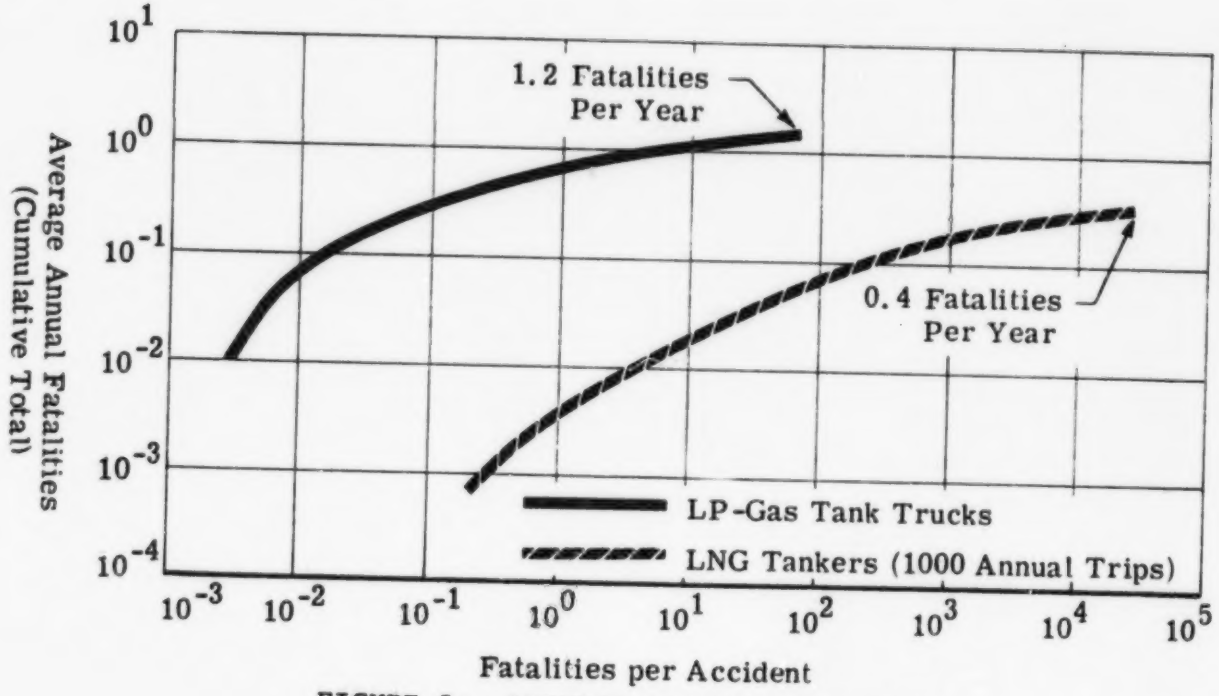


FIGURE 8: AVERAGE ANNUAL FATALITIES

For the years since 1961, an exact tally has not been made, but the annual average is believed to be in the range of 1 to 2 per year. The principal plume fire accident during this period was that killing 10 persons at Berlin, New York.[15]

## 6.0 Uncertainties

In spite of the good agreement obtained between the risk calculation for LP-Gas tank truck spills and data, there still exist considerable uncertainties in the calculation. These arise primarily in the probability values associated with the four variables.

Perhaps the greatest uncertainty is associated with the values of P(Q). For tank truck spills we have already noted that P(Q) is based on spill data that are known to be incomplete. The spill frequency based on accident reports was adjusted upward by a factor of 5 to reflect this. Hence the uncertainty in the value of P(Q) used could be as much as a factor of five.

For LNG tankship spills the values of P(Q) could be high by a factor of 100 which reflects the fact that no credit was given for the tankship's double hull and possible special navigational procedures. At the side walls, the two hulls are separated by several feet and are held together by many web frame members and stringers, giving a kind of honeycomb structure. For example, the probability of occurrence of an "undesirable event" to an LNG tanker has been estimated to be $1 \times 10^{-7}$ to $2 \times 10^{-7}$ per trip[28] compared with the value of $5 \times 10^{-5}$ (sum of leak and tank rupture incidents) used here.

The uncertainties regarding dispersion characteristics, V and P(V), are relatively small, less than a factor of two. The important item to note is that the values used are considered to be representative of the average of many sites. If the risk methodology were to be applied to a specific site, meteorology specific to that site should be

used, and in addition, consideration should be given to terrain features which also could affect dispersion. This latter is particularly true for sites along the Pacific coast where the land may rise steeply at or near the shore.

Uncertainties associated with I and P(I) are somewhat larger and are estimated to be in the range of a factor of 2 to 5. The principal contributor is the value of P(I) derived from reported accidents. The reports do not include all accidents, and those reported most likely are those with more severe consequence. Hence it is thought that the occurrence of smaller plumes is underestimated. This would lead to the result that the ignition probabilities for large areas are overestimated in Table XII. In this connection, it should be kept in mind that the ignition probabilities used for areas >30,000 $m^2$ represent an extrapolation of the reported data.

An offsetting uncertainty arises from the manner in which the plume areas were calculated. The values used for the dispersion coefficients, Table VII, are more appropriate for time-averaged plumes (10 minute averages). In contrast, the combustion of a flammable plume following ignition is relatively instantaneous. Generally the values of the dispersion coefficients used with short averaging times are smaller. This was ignored in these calculations, and as a result, the calculated areas for the steady spills may be too small by a factor of 2. For the reasons noted in Section 4.3.1 this effect should be of no consequence for the instantaneous spill plumes.

The values of $\rho$ and $P(\rho)$ were meant to be representative of all locations, collectively. Although they were derived in an approximate manner, their uncertainty for this use is estimated to be less than a factor of two for LP-Gas accidents and five for LNG accidents. However as with V and P(V), these values might be quite different by a larger factor for a specific site. Also to be kept in mind is that the population of a given area may vary significantly with the time of day. Finally, with regard to LNG tanker spills, the population density values were based on the Staten Island - New Jersey area. Other LNG terminal locations such as Cove Point, Maryland, and Point Concepcion, California, are rural in character with very low population densities. The use of the higher population densities to characterize LNG terminal sites is conservative and contributes to a higher estimated risk.

In summary, the overall uncertainty in the risk of fatalities from LNG tanker spills is estimated to be a factor of 2 to 1/100, mainly because of the disregard of the protection afforded by the tankship's double hull. For LP-Gas spills, the overall uncertainty is estimated to be a factor of 10 to 1/10.

## 7.0 Final Comments

Figure 9 compares the risk spectrum of LP-Gas tank truck and LNG tankship accidents with the spectrum of all fatal fires in the U.S. and Canada. The latter are based on data for fires and explosions during the years of 1965 to 1972 compiled annually by the National Fire Protection Association.[31] Data on large fires, $N \geq 100$, cover the years 1871 to 1972. These include the Chicago fire, 250 fatalities; the Iroquois Theater fire, 602 fatalities; the Cocoanut Grove night club fire, 491 fatalities; the munitions ship explosion in Halifax Harbor, 1,600 fatalities; and the Texas City ammonium nitrate fertilizer explosion and fire, 561 fatalities.

The estimated LP-Gas and LNG risk spectrums are less than one percent of all fires and explosions. However, these spectrums are for specific accident event sequences and do not include all sequences for all phases of LP-Gas and LNG storage and transport activities. Other fatality mechanisms and accident event chains need to be considered. For example, fatalities caused by overpressure effects, LNG tanker unloading operations, LNG transport via barge and tank truck, LP-Gas transport by pipeline and rail tank car need to be assessed. For such cases, direct application or minor extensions of this method will be adequate.

The fact that good agreement was obtained between the prediction of the risk assessment method and historical data for LP-Gas fatalities is promising. However, an important aspect requiring verification is the prediction of the frequency of the many levels of consequence. This may be

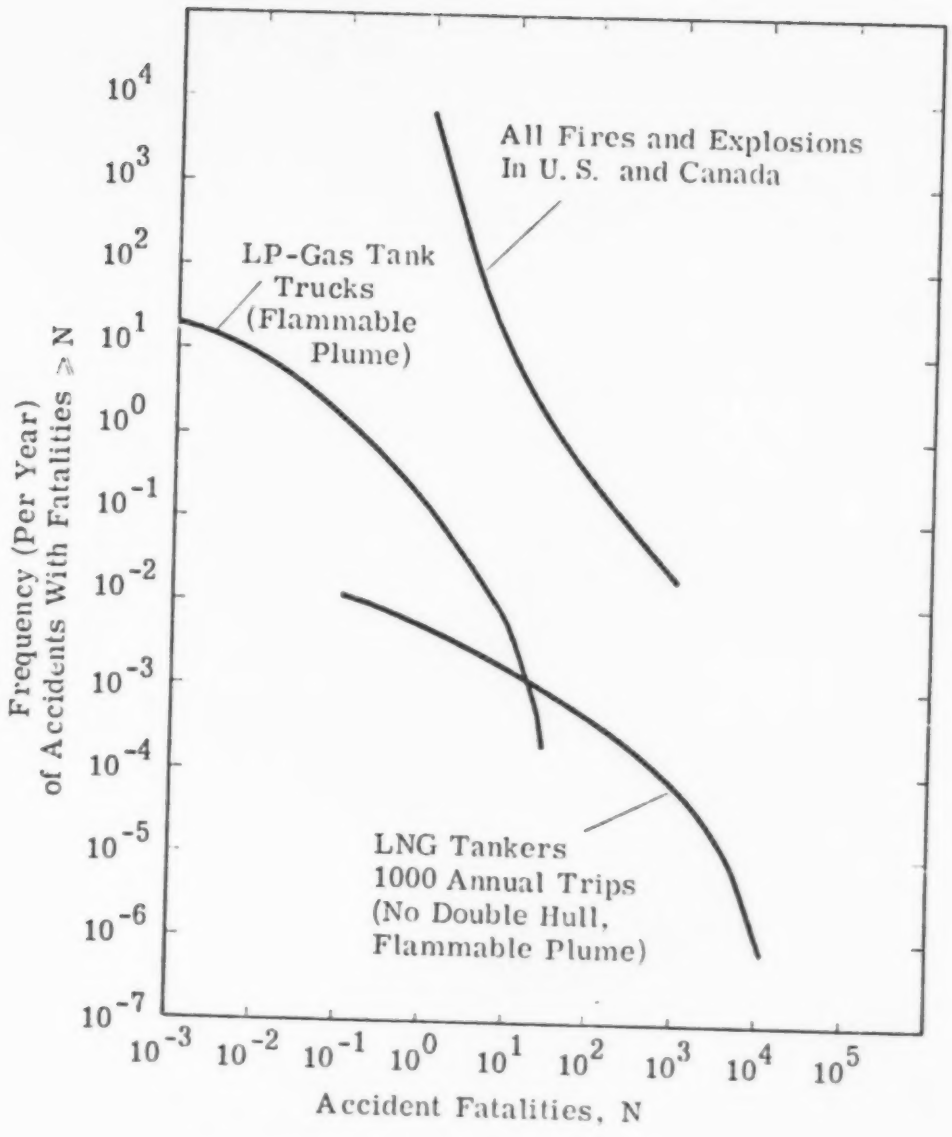


FIGURE 9. COMPARISON OF SOME FIRE RISKS

difficult to do, using fatality data only. The verification might be easier for assessments of injuries or property damage because of the greater frequency of accidents with these consequences.

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# APPENDIX A

(TABLE III)

TABLE III
SPILLS OF LP-GAS ASSOCIATED WITH
TANK TRUCK OPERATIONS: JULY 1971 - DECEMBER 1973

| Date, Place | Incident | Loss | Casualties |
|---|---|---|---|
| 7/8/71<br>Moran Junction<br>Wyoming | Roll-over of tank truck caused a leak in the rotary gauge. | 1,000 gal.<br>Propane | $13,500 damages |
| 8/3/71<br>Plattekill, N. Y. | Leak occurred during an unloading operation. Explosion. | 200 gal.<br>Propane | >$30,000 damages |
| 7/26/71<br>Thorofare, N. J. | Valve jarred open during a roll over of a tank truck. Valve closed to stop leak. | 100 gal.<br>Propylene | $11 damages |
| 8/27/71<br>Orangeburg Co.,<br>S.C. | Several minor leaks were caused by roll over of a tank truck. | 20 gal<br>Propane | $58 damages |
| 9/2/71<br>Platteville, Wis. | Transfer line ruptured during an unloading operation involving a 11,378 gal. tank truck. An explosion followed ignition by an auto. | 300 gal.<br>Propane | 1 injured<br>$6,900 damages |
| 11/17/71<br>Gillette, Wyoming | Transfer lines broken when driver pulled away without disconnecting them. Fire | 1000 gal.<br>Propane | 1 injured<br>$200 damages |
| 12/17/71<br>Horseheads, N. Y. | Driver drove away without disconnecting transfer lines. Propane leaked and ignited. | 25 gal.<br>Propane | $350 damages |
| 12/21/71<br>Providence, R.I. | Rotary gauge on a 10,400 gallon tank truck ruptured. No fire. | 1,200 gal.<br>Propane | $120 damages |

TABLE III (continued)

| Date, Place | Incident | Loss | Casualties |
|---|---|---|---|
| 1/8/72<br>Green River, Ut. | Tank truck overturned when ran off onto shoulder of road. Gauges broken off and two 5,000 gal. tanks punctured. | 9601 gal.<br>Propane | 1 killed<br>$10,146 damages |
| 1/10/72<br>Whitehall, N.Y. | Gauging devices leaked on 11,000 gal. tank. | 7 gal.<br>Propane | --- |
| 2/9/72<br>Tewksbury, Mass. | While unloading a tank truck, a complete break of an unloading pipe occurred at a common pump on the dock. Flash fire. | 14,805 gal.<br>Propane | 1 killed<br>23 injured |
| 3/9/72<br>Lynchburg, Va. | 10,600 gal. tank truck rolled over and the tank was punctured. An explosion resulted. | 9,200 gal.<br>Propane | 1 killed<br>6 injured<br>$100,000 damages |
| 2/18/72<br>Waldorf, Md. | 3/4-Inch pump bypass line broken while tank truck was being towed. | 1,500 gal.<br>Propane | Not reported |
| 5/14/72<br>Caro, Mich. | Small crack occurred in 10,970 gal. tank truck. Butylene leaked. | 366 gal.<br>Butylene | $30 damages |
| 6/2/72<br>Mendenhall, Miss. | Transfer hose ruptured and propane escaped from storage tank since the excess flow valve did not function. Storage tank exploded in the fire. | 30,000 gal.<br>Propane | 1 injured<br>$100,000 damages |
| 2/2/72<br>Van Nueys, Ca. | Tank truck rolled over causing a leak at a corrosion-weakened thermometer well. | 20 gal.<br>Butane | --- |
| 6/20/72<br>Chester, Va. | Driver drove off without disconnecting unloading line. No fire. | 1,700 gal.<br>Propane | $200 damages |

TABLE III (continued)

| Date, Place | Incident | Loss | Casualties |
|---|---|---|---|
| 7/14/72<br>Lancaster, Pa. | Pressure relief valve was broken as 10,190 gal. tank truck passed under a low bridge. | 2,500 gal<br>Propane | $350 damages |
| 7/14/72<br>Mt. Kisco, N.Y. | Flexible hose between 1,085 gal. and the pump on a truck failed. Flash fire. | 700 gal.<br>Propane | 4 injured<br>$15,000 damages |
| 7/27/72<br>Milwaukee, Wis. | Small leak occurred in line between tank and outlet shut-off valve. | Not known | --- |
| 8/8/72<br>Claremont, N.H. | Transfer line ruptured during unloading. Driver shut-off valves. No fire. | 50 gal.<br>Propane | $5 damages |
| 9/29/72<br>Northwood, Ia. | Vapor line broken when a bridge collapsed. No fire. | 2 gal.<br>Propane | $40,000 damages |
| 9/21/72<br>N.J. Turnpike,<br>N.J. | 8,000 gal. tank truck exploded following an accident. | 7,209 gal.<br>Propylene | 2 killed<br>1 injured<br>$31,000 damages |
| 10/10/72<br>Ridgefield, Conn. | Roll over caused a leak in a bleeder line. Fire. | 200 gal.<br>Propane | $2,000 damages |
| 10/17/72<br>Oakland, Ca. | Roll over caused a leak in the rotary gauge. | 300 gal.<br>Propane | 1 injured<br>$25,000 damages |
| 11/6/72<br>Ulysses, Kan. | Valve gasket failed during an unloading operation. | 280 gal.<br>Propane | $28 damages |
| 12/27/72<br>Tulsa, Okla. | Driver accidently opened wrong valve on tank truck. | 1 gal. | Driver received freeze burns |
| 12/28/72<br>Watkinsville, Ga. | Loss of control on curve caused accident resulting in failure of tank at a weld. | 8,500 gal.<br>Propane | Fire, $24,000 damages |

TABLE III (continued)

| Date, Place | Incident | Loss | Casualties |
|---|---|---|---|
| 2/22/73<br>Table Rock, Wyo. | Delivery tank truck hit by train at grade crossing resulting in puncture of the tank. | 2,500 gal.<br>Propane | 1 death<br>4 injured<br>$71,500 damages |
| 3/23/73<br>New Roads, La. | Hose ruptured while unloading | 600 gal.<br>Propane | Driver injured |
| 4/9/73<br>Greenville, S.C. | Delivery truck pulled away without disconnecting hose to 1,000 gal. storage tank. | 50 gal.<br>Propane | $50 damages |
| 5/17/73<br>St. Joe, Mo. | Tank truck collided with bridge and cracked unloading pipe. | 1,000 gal.<br>Propane | 1 injured<br>$24,000 damages |
| 5/28/73<br>Rocky Mount, N.C. | Fitting broke at unloading dock. Leaking propane ignited and the fire involved the tank truck (9,400 gal.) and two storage tanks (14,150 gal ea.) | 10,000 gal.<br>Propane | 1 injured<br>$75,000 damages |
| 6/12/73<br>West Grove, Pa. | Unloading hose broken when truck driver pulled away from unloading dock without disconnecting the hose. | 100 gal. | $500 damages |
| 5/12/73<br>Middletown, Pa. | At start of a delivery to a house, valve accidently opened before hose connection was made. | 1 gal.<br>Propane | 1 injured<br>(freeze burns) |
| 7/29/73<br>Boaz, Ala. | Worn hose on parked tank truck ruptured. | 300 gal.<br>Propane | $45 damages<br>No fire. |
| 8/21/73<br>Dover, Pa. | Tank truck pulled away from unloading dock without disconnecting hose. | 25 gal.<br>Propane | $1,000 damages |
| 9/12/73<br>Jacksonville, Fla. | Tank truck pulled away from unloading dock without disconnecting hose. | 25 gal.<br>Propane | $100 damages |
| 10/12/73<br>Moffit Co., Colo | Tank truck overturned on icy road. | 100 gal.<br>Butane | $25,000 damages |

TABLE III (continued)

| Date, Place | Incident | Loss | Casualties |
|---|---|---|---|
| 9/29/73<br>Portage, Ind. | As result of a rear-end collision, valve on tank truck was broken. | 225 gal.<br>Propane | 1 injured<br>$15,000 damages |
| 11/8/73<br>South Graham, N.C. | Pump valve broken off when truck rolled over. Driver closed off hand valve. | 5 gal.<br>Propane | ---- |
| 11/10/73<br>Orient, Ia. | Tank truck in collision with another vehicle. 2,600 gal. tank punctured. | 2,050 gal.<br>Propane | $16,390 damages |
| 11/20/73<br>Goshen, Va. | Tank truck fell off a weak bridge breaking a valve. | 8,700 gal.<br>Propane | $104,000 damages |
| 11/27/73<br>Russell, Mass. | Tank truck rolled over as the result of brake failure on a steep hill. Cause of leak not known. | 100 gal.<br>Propane | $24 damages |

# APPENDIX B

TABLE XI

ACCIDENTS INVOLVING SPILLS OF LP-GAS

AND OTHER FLAMMABLE LIQUIDS IN THE OPEN

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 7/20/1949 Woodbridge, N.Y. Commercial (a) | Plume from relief valve, 600 gal. tank being filled by tank truck. | Open flame hot water heater 65 ft. away | 80 | |
| 7/5/1938 George West, Tex. Restaurant, Residential (a) | Underground tank being filled from tank truck overflowed. Gas filled adjacent kitchen and house 50 ft. away. | Ignition in kitchen. | 41 | 4 killed, 20 injured. |
| 1/30/1939 DeGraff, Ohio Commercial (a) | Leak from underground tank being filled from cylinders. | "Nearby" coal stove in a restaurant. | <30 | 2 killed, 1 injured. |
| 6/13/1939 Brewster, N.Y. Commercial, Residential (a) | Broken valve on a LPG cylinder. | Stove burner in kitchen across 4-foot alley. | <30 | 3 killed. |
| 8/7/1946 Shreveport, La. Commercial (a) | Overfilled underground tank from a tank truck. | Restaurant stove inside but 7 1/2 feet from tank; violent explosion. | <30 | |
| 1/5/1952 Nashville, Ill. Commercial (a) | 50 gal. discharged from broken pipe extending from a truck; "knee deep" propane vapor cloud. | 40 ft. away in mercantile buildings. | <30 | |
| 4/25/1945 Los Angeles Factory (a) | Butane leaked from ruptured nipple at a condenser. Vapor spread to another building 150 ft. away. | At a boiler. | 370 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 5/2/1938 Detroit, Mich. Factory (a) | Butane leaked out of an unlighted furnace to another furnace 30 feet away. | In furnace. | <30 | |
| 10/13/1948 Sacramento, Ca. Railroad grade crossing (a) | Butane tank truck ruptured by a train collision at a grade crossing; released butane "enveloped area" in flames. | Not known but nearby. | <30 | 2 killed. |
| 6/11/1941 Los Angeles, Ca. Supply station for road grading job (a) | Butane discharged from broken filling lines for a 2,300 gal. tank truck. Butane "released over a wide area." | Acetylene torch. | 90 | Several injured in fire. |
| 1/18/1943 Los Angeles, Ca. Open road. (a) | Butane leaking from 4,300 gal. tank truck stopped at roadside. 1/4 Mile plume down road. | Ignited when taxi stopped opposite truck. | 10,000 | 5 killed in plume. |
| 11/21/1944 Denison, Tex. Street intersection (a) | 1000 gal. butane truck; nipple at bottom of tank broke; 600-foot diameter cloud. | Ignition 2 min. later at hot water heater 200 ft. away. | 1,500 | 10 killed in plume - liquid fire. |
| 8/23/1950 Wary, Colorado Commercial (a) | 2,782 gal. tank truck overturned breaking fill pipe; propane vapor cloud spread out. | In paint shop across the street. | 70 | 2 killed and 3 injured in flash fire. |
| 10/22/1936 Crowley, La. Service station (a) | Discharge pipe from 1,000 gal. tank containing 600 gal. of butane broke; vapor plume formed. | Ignited by trash burner 165 feet away. | 400 | 1 killed in fire flash. 4 killed in building (explosion). |
| 10/16/1941 Covington, Ky. Distribution Plant (a) | Butane gas leaked from unconnected vapor return line connected to truck but not the storage tank. | Ignition from back-fire or magneto contacts. | <30 | Flash fire injured 2. |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 7/18/1946 Big Bear Lake, Ca. Distribution Plant (a) | Butane line, loosened by previous earthquake, broke and vapor cloud spread in all directions over a two-acre area. | Ignition occurred when cloud reached one of the nearby cabins. | 8,200 | 1 injured by flash fire. |
| 9/29/1947 Spencer, W. Va. Distribution plant in town (a) | Fitting on skid tank broke in cylinder loading room. Gas flowed down incline toward dwellings. | Ignited in a house 130 ft. away. Explosions in other dwellings, plant and commercial buildings. | 275 | |
| 2/8/1951 St. Elmo. Ill. Distribution plant (a) | 9,500 gal. of butane escaped from a broken pipe to a 14,000 gal. above-ground storage tank. Excess flow valve did not function. | Did not ignite. Ignition sources removed and cloud dissipated. Cold - ice and snow. A train passed through part of the cloud. | - | |
| 11/19/1946 Greenville, S.C. Commercial, Laundry yard (a) | Recently installed 6,000 gal. propane tank developed a leak which could not be closed off. Vapor cloud formed for 27 min. in early evening. | Ignition source unknown. Explosion destroyed all buildings within a radius of 300 ft. | 1,500 | 6 people killed, 150 injured. |
| 10/18/1948 Texas City Commerical - Industrial in early evening (a) | Pipeline break (propane - propylene) created vapor-air cloud covering 68,000 $ft^2$ across highway. | Several autos stalled in cloud before ignition occurred. | 6,300 | 7 killed, 14 injured. |
| 7/21/1952 Bakersfield, Ca. Petroleum Refinery (b) | Earthquake toppled butane storage tank breaking connecting piping. Vapor air cloud covered 5 acres before igniting. | Believed to have occured in electrical equipment | 21,000 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area($m^2$) | Casualties |
|---|---|---|---|---|
| 1/8/1957 Montreal East,Que. Petroleum Refinery(b) | Butane storage sphere was over filled and liquid collected in diked area. Undetected, a large vapor cloud formed during the night. | Ignition in a service station 600 ft. away. | 5,900 | |
| 2/15/1958 Alam, Mich. Petroleum Refinery (b) | Overpressurized tank ruptured and formed a visible fog. | Ignited when reached power house 350 ft. away. | 2,000 | |
| 5/28/1959 McKittrick Petroleum Refinery (b) | 1-Inch pipe connection on a LP-Gas storage tank broke. Vapor air cloud formed. | Ignited by gas fueled engine 50 ft. away. | 41 | |
| 4/10/1954 Fredericksburg, Va. Gas distribution and storage plant (b) | 2-Inch pipe beneath two 30,000 gal. storage tanks broke. | The leaking gas was ignited by an open flame 6 ft. away. | <30 | |
| 10/14/1955 Elkhart, Ind. Commercial (b) | 1 1/2-Inch LP-Gas line was broken at a service station at 5:30 A.M. by a truck. Low hanging cloud spread over adjacent highway and into a paint shop. | Ignition by a truck driver who stopped in the "foggy" cloud. | 160 | 2 killed. |
| 12/14/1955 Moopa, Ca. Gas distribution and storage (b) | Truck loading line broke when driver failed to disconnect hose and drove away. LP vapor-air cloud 3 feet deep, covering 300 $yd^2$ formed. | Ignition by pilot lights in dwellings. | 250 | |
| 9/4/1957 Jackson, Miss. Commercial (b) | LP-Gas truck-to-storage tank transfer line broke. Gas cloud spread to across street where a workman's auto was parked. | Ignition when workman attempted to start auto and escape. | 160 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 3/12/1959 Brinkely, Ark. Commercial | LP-Gas truck driver bent the gas manifold and broke the unloading line while moving his truck during filling of a storage tank. | Ignition by truck engine. | <30 | |
| 10/18/1954 Portland, Ore. Rail yard. (b) | LP-Gas tank car observed to leak; vapor cloud traveled 200 feet. | Ignition at kerosine switch lamp. Explosive violence reached after flame traveled a few feet. | 660 | |
| 7/24/1952 Kansas City, Kan. Service station (b) | LP-Gas fuel tank over filled on hot day. Relief valve functioned and plume entered station office. | Ignition at gas refrigerator. | <30 | |
| 10/5/1956 Carthage, Tex. Highway (b) | Driver stopped to bleed pressure from 5,900 gal. tank trailer. Gas cloud formed. | Ignited when a parked auto was started 30 feet away. | <30 | 4 killed, 2 injured. |
| 10/18/1956 Herrin, Ill. Residential - Commercial (b) | Leak in tank truck filling line allowed gas cloud to spread into residential area. ∿1,000 gal. leaked away. | Ignited explosively in dwelling 175 feet away. | 500 | 9 killed. |
| 6/23/1957 Sylvania, Ga. Commercial (b) | Leak from flexible hose connecting cargo tank to pump. Propane-air cloud formed. | Ignited explosively in a restaurant kitchen 65 ft. away. | 69 | |
| 12/21/1957 Southwide, Mass. Residential (b) | Nozzle on a hose connecting supply truck and outside, 100-gal cylinder broke. Gas seeped into house through open door. | Ignited at kitchen stove. | <30 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 11/14/1958 Los Angeles, Ca. Highway (b) | 3/4-Inch plug in guaging hole knocked out when tank truck overturned on highway. LP-Gas cloud formed and spread across 3-lane road and into storm sewer opening. Leak subsequently stopped up. | No ignition. | - | |
| 7/2/1958 Charleston, S.C. Rural (c) | Gas transmission line broken by earth-moving machinery. | Ignition was immediate. | <30 | |
| 3/6/1951 Kobuta, Pa. Butadiene Plant (d) | Butadiene accidentally discharged through an open 1/2-inch valve. | Ignited by an open flame heater 185 ft. away. | 560 | |
| 7/15/1960 Fort Devens, Mass. Army Post (e) | 2-Inch transfer line connection broke during unloading operation. Vapor air plume formed (LP-gas). | Ignition 25 ft. away in mess hall kitchen. | <30 | |
| 7/25/1962 Berlin, N. Y. Rural (f) | Semi-trailer, LP-Gas tank over-turned rounding a curve, rupturing the tank and releasing 7,000 gal. of LPG. Vapor-air cloud formed covering about 200,000 $ft^2$ - 80 ft high. | Ignition source unknown but delayed until nearly all LPG vaporized - a few minutes - several buildings infiltrated with gas. | 19,000 | 10 killed, 17 injured. |
| 6/2/1959 Dear Lake, Pa. Highway (g) | LP-Gas truck struck from behind at traffic light. | Immediate ignition. | <30 | |
| 7/5/1973 Kingman, Arz. LP-Gas Distribution (h) | Unloading line started to leak while setting up for unloading LP-Gas tank car. | Immediate ignition. | <30 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 1972<br>Hopkinton, Mass.<br>Highway (i) | LP-Gas tank truck punctured in accident. | Immediate ignition. | <30 | |
| 12/9/1970<br>Port Hudson, Mo.<br>Rural (k) | LP-Gas pipeline ruptured at about 10:00 P.M. in the evening, discharging about 740 barrels (31 bbls/min) before ignition occurred. Vapor air cloud spread down-slope and down-wind about 1,000 ft. before reaching some farm buildings. | Ignition occurred in one of the farm buildings | 16,000 | |
| 2/22/1973<br>Austin, Tex.<br>Rural (l) | 10-Inch natural gas liquids pipeline broke and a few thousand barrels escaped. The liquid flowed out of a pump station area and into ditches along a road for 2,400 ft. Two cars stalled in the two-foot thick vapor air cloud. | A third car entered the cloud and stalled. Ignition occurred when attempted restart 2,000 ft. from pipeline break. | 7,500 | 6 killed, 2 injured. |
| 1/22/1972<br>East St. Louis<br>Rail yard (m) | During a "humping" operation a tank car containing liquid propylene collided with another car and was punctured. The resulting vapor air cloud covered 5 acres before ignition. | Ignition near an un-occupied caboose. | 21,000 | 240 injured. |
| 10/19/1971<br>Houston, Tex.<br>Railroad (j) | Train derailed and two tank cars of vinyl chloride ruptured. | Immediate ignition. | <30 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 7/31/1963 Memphis, Tenn. Residential (n) | 7,000 gal. LP-Gas tank truck hose broke and leaked LPG while in transit. Leak observed and truck stopped. | Ignition source not known but within 20 ft. of truck. | <30 | 1 killed, 1 injured. |
| 6/16/1966 New York Harbor (o) | Texaco Massachusetts collided with and holed the number 1 tank to a depth of 12 feet on the Alva Cape. Texaco Massachusetts backed away; naptha gushed from the holed tank and spread out over the water between the two ships. | Ignition of the vapor air cloud occurred 3 min. min. after the accident. The Source was the tug Latin American, located between | 940 | 33 killed. |
| 12/1/1961 Luthersburg, Pa. Rural (b) | 12-Inch natural gas pipeline ruptured at a pumping station. | Immediate ignition. | <30 | |
| 6/28/1959 Meldrim, Ga. Rural (p) | LP-Gas tank car punctured during a derailment adjacent to a picnic ground. Vapor air cloud covered much of the picnic ground. | Ignition by a cooking fire estimated to be within 200 ft. | 650 | 23 killed. |
| 3/3/1956 San Jose, Ca. Commercial, (b) | While butane was being transferred from a tank truck to an above-ground storage tank, the transfer hose developed a leak. | Vapors ignited by truck ignition system. | <30 | 1 injured. |
| 12/9/1959 Sacramento, Ca. Commercial (b) | LP-Gas tank fell off of a fork lift truck, breaking the relief valve. | Immediate ignition by the engine of the fork lift truck. | <30 | |

TABLE XI (continued)

| Date, Location and Reference | Accident | Ignition | Plume Area ($m^2$) | Casualties |
|---|---|---|---|---|
| 4/27/1961 Sidney, Neb. Highway (b) | LP-Gas truck overturned rounding a curve and a valve was broken off. | Escaping gas was ignited immediately. | <30 | 1 injured. |
| 12/29/1960 Annandale, Va. Residential (b) | During transfer of LP-Gas from a tank truck to a dwelling storage tank, the transfer line ruptured. | Immediate ignition. | <30 | |
| 12/21/1960 Los Angeles, Ca. Commercial (b) | Valve on a 388 gal. LP-Gas tank was broken off releasing LPG while the tank was being hoisted off the back of a flat-bed truck. | Ignition was in the immediate vicinity of the truck. | <30 | 2 killed. |
| 7/14/1959 Max Meadows, Va. LP-Gas Distribution Plant (b) | Transfer hose broke while a tank truck was being filled with LP-Gas from an above-ground storage tank. | The released vapor ignited near the tank truck. | <30 | |
| 10/11/1945 Laredo, Tex. Refinery (a) | Transfer line between a tank truck and a storage tank leaked. | Ignition occurred via the truck's engine. | <30 | |
| 10/7/1950 Woodbury, Ga. Commercial (a) | Transfer line between LP-Gas tank truck and one of six 1,000 gal. storage tanks broke at the tank. | Vapors ignited at one of the storage tanks. | <30 | |

TABLE XI (continued)

FOOTNOTES

a. "LP-Gas Fires and Explosions", NFPA Quart 46, 46-82 (1952).

b. "LP-Gas Fires and Explosions", NFPA Quart 55, 169-212 (1961).

c. NFPA Quart 52, 320 (1959).

d. Cullivane, W. F.; G. F. Prince, and E. J. Dean, "Butadiene Fire, Kobuta, Pa." NFPA Quart 45, 66 (1951).

e. Jones. C. F., "LP-Gas Tank Truck Fire at Fort Devens", NFPA Quart 54, 227 (1961).

f. Walls, W. L., "LP-Gas Tank Truck Accident and Fire, Berlin, New York", NFPA Quart 57, 9 (1963).

g. Jones, C.F., "Fatal LP-Gas Tank Truck Fire", NFPA Quart. 53, 89 (1959).

h. Sharry, J. A., and W. L. Walls, "LP-Gas Distribution Plant Fire", Fire J. 68, 52 (1973).

i.

j. Hayes, T.C., Fire J. 66, 13 (1972).

k. Burgess, D. S. and M. G. Zabetakis, "Detonation of a Flammable Cloud Following a Propane Pipeline Break", Bureau of Mines Report of Investigations 7752 (1973).

l. "Pipeline Accident Report, Phillips Pipe Line Company Natural Gas Liquids Fire, Austin, Texas, February 22, 1973", National Transportation Safety Board Report No. NTSB-PAR-73-4 (1973).

m. "Railroad Accident Report, Hazardous Materials Railroad Accident in the Alton and Southern Gateway Yard in East St. Louis, Ill., January 22, 1972", National Transportation Safety Board Report No. NTSB-PAR-73-1 (1973).

n. "LP-Gas Transport Truck Fire", NFPA Quart 57, 406 (1964).

o. "Collision SS TEXACO MASSACHUSETTS and MV ALVA CAPE, June 16, 1966", National Transportation Safety Board Marine Accident Report, October 24, 1967.

p. "Large Loss of Life Fires - Transportation", NFPA Quart 54, 35-36 (1960).